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Industrial Wastewater Reclamation with a 400,000-Gallon-per-Day Vertical Tube Evaporator

General Tire and Rubber Co, Akron, Ohio

Prepared for

Industrial Environmental Research Lab -Cincinnati, Ohio

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16. ABSTRACT

A vertical tube evaporator (VTE) was built to recover water from a synthetic rubber manufacturing plant wastewater stream containing 3,500 ppm dissolved solids, mostly chlorides and sulfates, and organics in excess of 100 ppm. Unit designed to produce 10.5 lbs. of water per pound of steam. Recovered water, containing near-zero organics and solids, was recycled to the manufacturing process. Performance over short periods exceeded design. Continuous operation for extended periods was rendered impossible, due to fouling and corrosion of copper alloy heat exchange surfaces. Corrosion was traced to presence of 2 to 5 ppm of sulfides in the wastewater feed. Unit was retubed with titanium. This eliminated the corrosion, but the fouling continued. All attempts to reduce fouling by pretreating the wastewater feed stream were unsuccessful. Outside of the problems caused by the fouling of the heat transfer surfaces, the unit operated satisfactorily. If the fouling can be brought under control, this method has a high probability of providing an economical method of renovating wastewater streams containing water soluble salts and organic chemicals. This report was submitted in fulfillment of R & D Project 12020 GUT, Program Element IBB610, by The General Tire & Rubber Company under the partial sponsorship of the Environmental Protection Agency. Work completed January 1975.

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INDUSTRIAL WASTEWATER RECLAMATION WITH A 400,000-GALLON-PER-DAY VERTICAL TUBE EVAPORATOR

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FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report, "Industrial Wastewater Reclamation with a 400,000 Gallon-Per Day Vertical Tube Evaporator", is a product of the above efforts. It concludes that physical/chemical treatment methods continue to be a most important tool in the control of industrial pollution. This study combined several program objectives in demonstrating one such treatment method, multi-stage evaporation, for pollution control and the associated reuse of the recovered water in the process. These results can be applied by the rubber manufacturing industry, and others, in their search for more complete water reuse. For further information on the subject, contact the Industrial Environmental Research Laboratory, Edison, NJ field station 08817.

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ABSTRACT

A 14-effect vertical tube evaporator (VTE) was built to recover water from a synthetic rubber manufacturing plant wastewater stream containing a 3,500 ppm dissolved solids, mostly chlorides and sulfates, and organics in excess of 100 ppm. The unit was designed to produce 10.5 pounds of water per pound of steam. Recovered water, containing near-zero organics and very low total dissolved solids, was recycled to the manufacturing process.

Performance over short periods exceeded design. Continuous operation for extended periods was rendered impossible due to fouling and corrosion of the copper alloy heat exchange surfaces. Corrosion was traced to the presence of 2 to 5 ppm of sulfides in the wastewater feed.

The unit was retubed with titanium. This eliminated the corrosion, but the fouling continued. All attempts to reduce fouling by pretreating the wastewater feed stream were unsuccessful.

Except for the problems caused by the fouling of the heat transfer surfaces, the unit operated satisfactorily. If the fouling can be brought under control, this method has a high probability of providing an economical method of renovating wastewater streams containing water soluble salts and organic chemicals. Direct operating costs varied from $0.80 to $8.44 per thousand gallons of water recovered.

This report was submitted in fulfillment of R & D Project 12020 GUT by The General Tire & Rubber Company under the partial sponsorship of the U.S. Environmental Protection Agency. The work was completed in January 1975.

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INTRODUCTION

This report covers the installation and operation of a 14-effect evaporator located at The General Tire & Rubber Company's synthetic rubber plant at Odessa, Texas. This plant has a nominal rated capacity of 40,000 tons per year of butadiene-styrene type synthetic rubber (GRS) and employs 250 people. It is located in a semi-arid area of Western Texas, and there are no permanent flowing streams in the vicinity to provide dilution of the plant effluent. This plant was constructed in 1957. For the first 10 years, the plant effluent was used by an oil producing company for water flooding of an oil field for the secondary recovery of crude oil. At the completion of the 10-year contract period, the oil company declined to exercise their option for renewal due to the difficulties they had encountered in treating the water for water flooding which is the same as would be required for the deep well injection method of disposal. In 1968, an 89-acre plastic lined solar evaporation pond was constructed as a temporary measure and also as the ultimate final sink for the disposal of the concentrated brine which would be the by-product of any process to recover and reuse part of the water.

As a permanent solution to the disposal problem, it will be necessary to recover part of the water in order to reduce the volume of brine to be disposed of by solar evaporation, deep well injection or spray drying. The use of membrane-type processes such as reverse osmosis and electrical dialysis for the recovery of the wastewater were considered to have a much lesser chance of success than the distillation process due to the fact that any fouling of the membranes would seriously reduce or completely eliminate the possibility of producing high-quality product water whereas with the distillation unit fouling of the heat transfer surface would only reduce the efficiency of the unit and would not seriously hamper its ability to produce high-quality reusable water at a reasonable rate.

The economical removal of water soluble salts from plant effluents has always caused an almost insurmountable problem. Although the production of distilled water from brines has been accomplished for a considerable time, the end use of the water on the larger units was for drinking purposes; and as generally there was no other fresh water in the area, cost was of no consideration. It should also be noted that these desalination plants used as a feed either well water containing a high salt content or sea water. In either case, the feed stock had more or less a constant composition and was free from organic contaminants. In treating industrial effluents in addition to the dissolved salts, organic chemicals may also be present, and the composition of the effluent is not constant as it must accept whatever wastes the production plant wishes to dispose of. The conceptual design of the 14-stage vertical tube evaporator erected at the synthetic rubber plant, owned and operated by The General Tire & Rubber Company, at Odessa, Texas, indicated that the effluent from the plant could be treated in an evaporator and would accomplish three goals:

1. Recover up to a maximum of 80% of the water for reuse in the plant, thus alleviating a possible water shortage in case of a serious drought in this semi-arid area.
2. Reduce the quantity of concentrated brine that would have to be disposed of in the solar evaporation ponds.
3. Accomplish the above at a cost of between $1 and $1.25 per thousand gallons of water reclaimed, including the charges for depreciation of the capital equipment required.

The facility built under this grant demonstrates the applicability of a multiple effect vertical tube evaporator to the renovation of chemical plant wastes containing both inorganic and organic chemicals. The product water contained less than 10 parts per million of dissolved

salts, and the volume of the concentrated brine was reduced to a reasonable volume for ultimate disposal. While this process was demonstrated on the effluent from a synthetic rubber plant producing GRS type rubber, the principle is applicable to a wide spectrum of chemical and industrial plant effluents, including those where the ultimate disposal of the brine requires it to be taken to dryness in an auxiliary piece of equipment such as a flash dryer.

The chemical wastewater effluent from The General Tire & Rubber Company's Synthetic Rubber Plant, Odessa, Texas, at flow rates up to 750,000 gpd contains dissolved solids, mostly sulfates and chlorides in concentrations of approximately 3,500 ppm, in addition to organics of approximately 250 ppm. The evaporation plant is used to obtain high-quality water for reuse in the synthetic rubber manufacturing process. The residual concentrated brine is disposed of by means of existing plastic-lined solar evaporation ponds containing a total area of 89 acres.

In order to prevent ground water pollution, for the past several years, The General Tire & Rubber Company Odessa Plant had been disposing its aqueous industrial wastes by discharging them into these solar evaporation ponds. These ponds gradually filled to the point where it was probable that production would have had to be curtailed within a short time. An alternate solution to waste handling was therefore essential. From theoretical and pilot plant studies, The General Tire & Rubber Company decided that evaporation of the rubber plant wastewater by the vertical tube process coupled with discharge of the concentrated brine to the existing solar evaporation ponds would provide wastewater renovation and an ultimate wastewater disposal system that completely abated water pollution and also would be economically acceptable.

Specific objectives of the project were to determine:

1. The ratio of the quantity of pure product water to the quantity of wastewater feed.

2. The amount of pure product water per unit of thermal energy applied, the "gain ratio" or "economy ratio."

3. The performance of the plant at various feed rates.

4. The corrosion resistance of the construction materials.

5. The rate of scaling or fouling, if any, on the heat exchange surfaces.

6. The overall utilities requirements.

7. The engineering data to further optimize the design of future VTE plants for industrial wastewater application.

8. The concentrations of organics and dissolved solids in the pure product water compared to concentrations in the wastewater feed.

9. From the above, the cost of the reclaimed water.

The engineering design of the VTE facility, materials procurement, and start-up services were furnished by Envirogenics Company, El Monte, California. Engineering design of auxiliary facilities and construction supervision was carried out by the Central Engineering Department of The General Tire & Rubber Company, Akron, Ohio.

After a brief initial period of operation at above design specifications, the VTE performance rapidly deteriorated, due to fouling and corrosion of the copper-nickel alloy feedheater and evaporator tubes. The focus of development efforts throughout the balance of the demonstration period was on identifying the cause of and evaluating remedies for the fouling/corrosion problem.

CONCLUSIONS

The applicability of a 14-effect vertical tube evaporator (referred to as VTE) for the purpose of renovating an industrial waste stream containing approximately 3,500 ppm of inorganic salts, mostly sodium and potassium chlorides and sulfates, and approximately 100 ppm of organic chemicals plus 200 ppm of ammonium ion was demonstrated for a period of approximately two years with the following results:

1. The product water was uniformly of very high quality, containing less than 10 ppm of total dissolved solids; ammonia was reduced to one-fifteenth (1/15) of the concentration in the feed stock to the VTE. In order to reduce the ammonia to 3 to 5 ppm as required by the plant, the ammonia in the raw effluent was reduced from 200 ppm to 50 ppm by process changes elsewhere in the plant. The water was successfully used as process water throughout the demonstration period in a GRS type synthetic rubber plant at a rate of 200,000 to 300,000 gallons per day.

2. As a result of the removal of the high quality water for reuse, the volume of the brine containing the water soluble solids was reduced to approximately 35% of its original volume. Because of the fouling problems encountered, the unit was deliberately run at a low recovery rate. The recovery rate is governed only by the fouling characteristics of the effluent feed.

3. High unexpected corrosion rates on the 70-30 and 90-10 copper-nickel alloys used for the heat transfer surfaces was encountered. This was traced to the presence of 2 to 5 ppm of sulfide ion in the effluent feed stock. These sulfides resulted in part from the anaerobic bacteriological decomposition of the sulfates in the feed holding tanks and partly to sulfur - containing organic chemicals used in the plant.

4. In addition to the fouling caused by the sulfide corrosion, serious fouling also resulted from the organic chemicals and precipitation of iron and aluminum compounds.

5. Retubing the heat exchangers and evaporator with titanium tubes eliminated the corrosion problem but not all of the fouling problems.

6. The high rate of fouling resulted in a typical operating cycle of three (3) weeks on stream and one week down for cleaning which caused high maintenance and operating costs. These direct costs varied from $0.80 per thousand gallons to $8.44 (see Table 13).

7. Calcium sulfate scaling did not occur.

RECOMMENDATIONS AND FUTURE WORK

Although the product water was very pure, less than 10 parts per million of dissolved solids, and the unit operated unattended while reclaiming wastewater from the butadiene-styrene synthetic rubber plant, to evaluate this type of equipment for other processes, pilot scale experiments are necessary using actual virgin plant effluents in order to accurately assess such parameters as corrosivity and fouling tendencies. The economical use of this type of equipment depends on maintaining continuously a high heat transfer coefficient and having no corrosion or fouling of the thin-wall heat transfer surfaces.

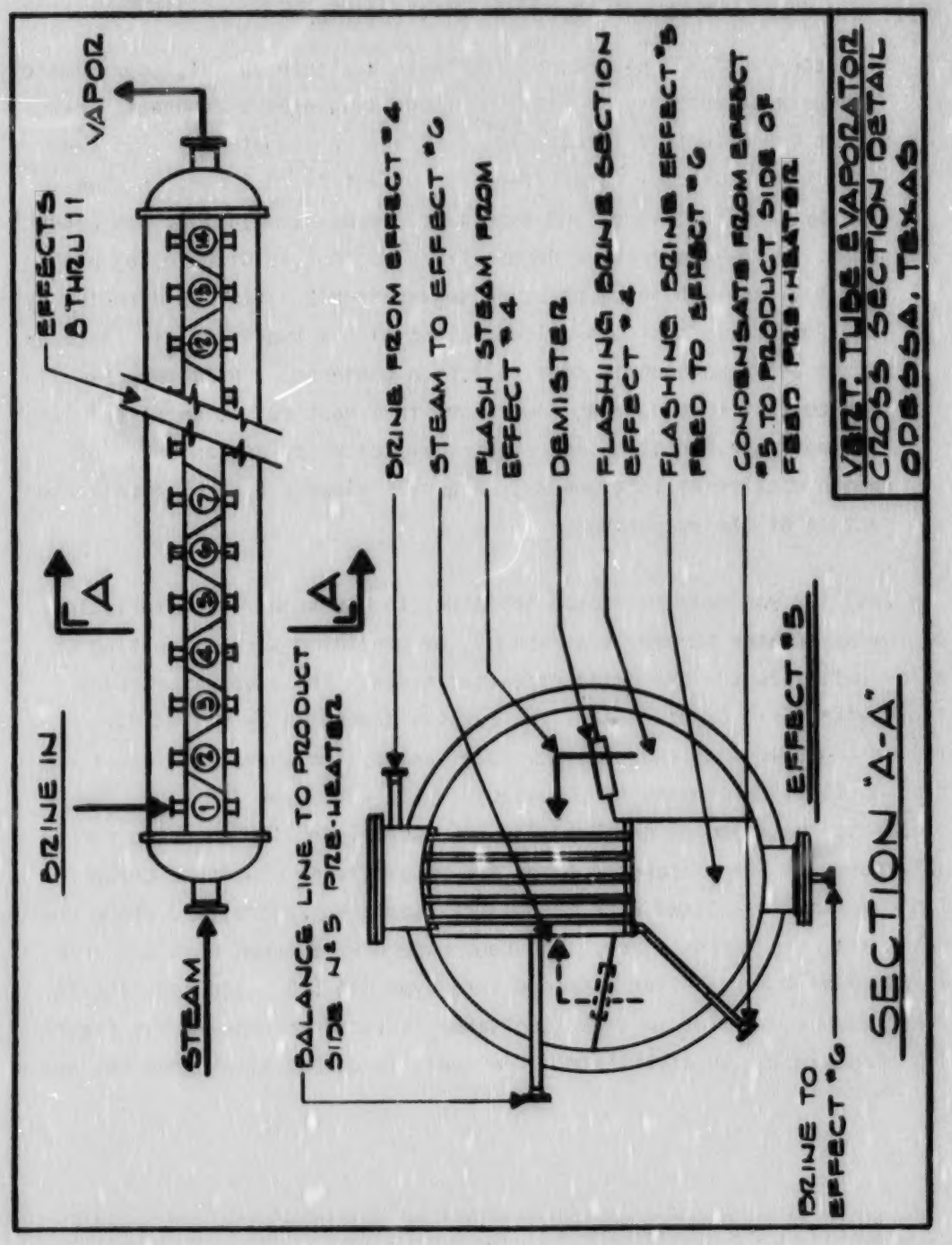
MECHANICAL DESCRIPTION OF THE VTE

Major components of the VTE as originally built are:

Fourteen vertical evaporator effects in a single vessel, approximately 9 feet in diameter by 73 feet long; one single-pass feedheater per effect for effects 1 through 13, 2.0 feet in diameter by 17.5 feet long; two, four-pass feedheaters for effect 14, 3.0 feet in diameter by 17.5 feet long; and associated feed, circulating, and product pumps. A low-temperature degassifier 3.5 feet in diameter by 36.3 feet tall and a high-temperature degassifier 2.0 feet in diameter by 13.0 feet tall is also provided; along with a two-stage air ejection system and a barometric heat rejection condenser. Auxiliary facilities consist of a cooling tower for final heat rejection, a building housing instrumentation and electrical controls, and an array of remote monitering instruments. Figure 1 shows a typical transverse section of the evaporator.

The facility was made corrosion resistant to the known characteristics of the wastewater stream (see Table 1) by providing surface coating or alloy materials for the brine-contacted areas. The evaporators were tubed with 90-10 Cu-Ni tubes 2 in. outside dimension by 6 feet long, roller expanded into carbon steel tube sheets, the upper tube sheet was clad on the side exposed to the liquid waste with Type 316 S.S. The evaporator water boxes are Type 316 S.S. with Type 316 S.S. clad removable covers. The evaporator sumps are mono-calcium aluminate cement-lined up to the knitted wire mesh entrainment separators, and above that point they are carbon steel. The feedheaters were tubed with 3/4 inch 70-30 Cu-Ni tubes, roller expanded into Type 316 S.S. clad tube sheets. The condensate outlet in each feedheater is sized to receive the flashing accumulation of distillate. The shell is carbon steel, but the heads

Figure 1. Vertical tube evaporator cross section detail



are Type 316 S.S. All feedheater and evaporator tubes are accessible for mechanical cleaning or replacement, if required.

Pumps are all Type 316 S.S. insofar as wetted parts are concerned, except distillate, cooling water, and condensate pumps which are cast iron, with bronze or stainless steel trim, as is appropriate. Pipelines are glass-reinforced epoxy or stainless steel for handling waste and concentrate; and carbon steel for water and steam vapor. The degassifiers are epoxy lined.

Both the feedheater and evaporator tubes are "enhanced." The feedheater tubes are spirally indented and the evaporator tubes are longitudinally corrugated. Design coefficients for feedheaters were developed by extensive tests at the University of Michigan, Heat Transfer Research Institute.[1] The design heat transfer coefficient values used vary with absolute temperature and range from a low of 620 to a high of 1010 BTU per hour-sq. ft.-F^{o}. Evaporator tubes of the longitudinally corrugated type were installed in the Envirogenics test evaporator, which was then fed actual wastewater from the Odessa Plant and operated over the bulk of the plant design conditions. The resulting coefficients were used for design purposes. They also vary with absolute temperature and have values ranging from a low of 660 to a high of 1,170 BTU per hour-sq. ft.-F^{o}. The final design was augmented by providing approximately 10% excess tubes over the design value.

As discussed elsewhere in this report, sulfide attack on the 90-10 and 70-30 copper-nickel heat exchange surfaces led to the decision to replace both feedheater and effect tubing with titanium tubes of similar enhanced surface characteristics.

PROCESS DESIGN AND DESCRIPTION OF THE VTE

SUMMARY

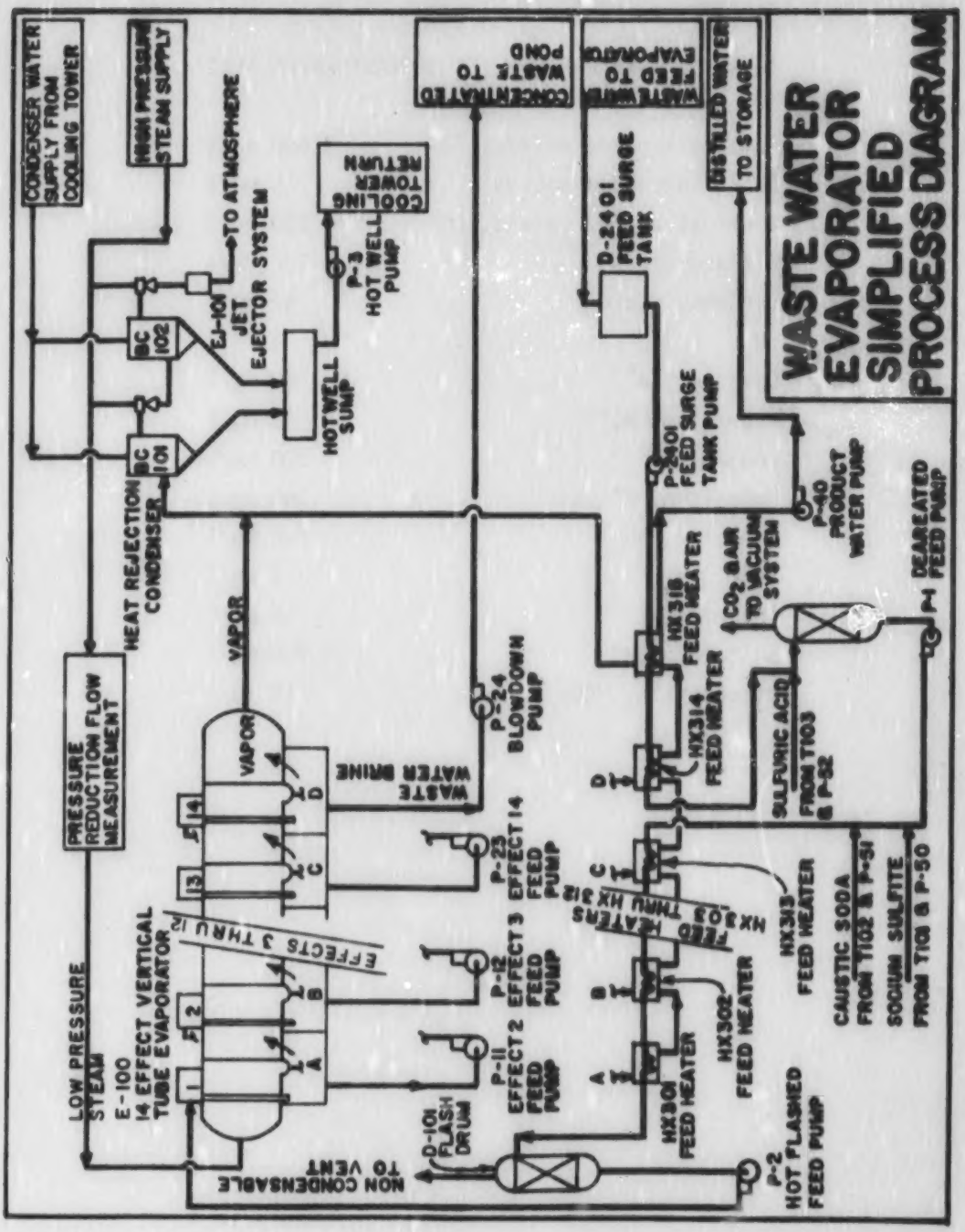
The plant is a 14-effect vertical tube evaporator, forward fed, and regeneratively feed-heated in the same number of stages as the effects. It is designed to receive clarified wastewater normally containing 3,500 ppm total dissolved solids at a pH of 6 and 75° F. This feed can be concentrated to 6.7 times its inlet concentration, by removing 85% of its weight as recovered water which is delivered from the sump of the last feedheater at 110° F. The concentrate is discharged from the sump of effect 14 at 110.4° F. The concentration limit is governed by the scaling characteristics of the feedwater.

A simplified process flow diagram is shown on Figure 2.

WASTEWATER CHARACTERISTICS

Separate wastewater streams emanate from the major operations of the synthetic rubber plant and are combined in a neutralization pit and clarified prior to discharge to a holding pond. Sanitary wastes are treated in the municipal sewage treatment plant. The present rubber plant total wastewater effluent flow rate averages 0.523 million gallons per day (mgd) and is expected to increase to 0.74 mgd upon anticipated plant expansion.

Figure 2. Simplified Process Flow Diagram


WASTE WATER EVAPORATOR SIMPLIFIED PROCESS DIAGRAM
CONDENSER WATER SUPPLY FROM COOLING TOWER
HIGH PRESSURE STEAM SUPPLY
TO ATMOSPHERE
EJ-101 JET EJECTOR SYSTEM
BC 102
BC 101
HOTWELL SUMP
P-3 HOT WELL PUMP
COOLING TOWER RETURN
CONCENTRATED WASTE TO POND
WASTE FEED TO EVAPORATOR
D-2401 FEED SURGE TANK
DISTILLED WATER TO STORAGE
P-2401 FEED SURGE TANK PUMP
P-40 PRODUCT WATER PUMP
DEAREATED FEED PUMP
HEAT REJECTION CONDENSER
PRESSURE REDUCTION FLOW MEASUREMENT
VAPOR
WASTE WATER BRINE
P-24 BLOWDOWN PUMP
HX316 FEED HEATER
HX314 FEED HEATER
CO_2 & AIR TO VACUUM SYSTEM
SULFURIC ACID FROM T103 & P-52
LOW PRESSURE STEAM
E-100 14 EFFECT VERTICAL TUBE EVAPORATOR
EFFECTS 3 THRU 12
EFFECTS 13 & 14
P-23 EFFECT 14 FEED PUMP
P-12 EFFECT 3 FEED PUMP
P-11 EFFECT 2 FEED PUMP
FEED HEATERS 3 THRU HX 312
HX313 FEED HEATER
HX302 FEED HEATER
HX301 FEED HEATER
CAUSTIC SODA FROM T102 & P-51
SODIUM SULFITE FROM T101 & P-50
D-101 FLASH DRUM
NON CONDENSABLE TO VENT
P-2 HOT FLASHED FEED PUMP

The feed to the VTE unit was taken from this holding pond. The typical characteristics of this wastewater are presented in Table 1.

Table 1. TYPICAL QUALITY OF WASTEWATER FEED

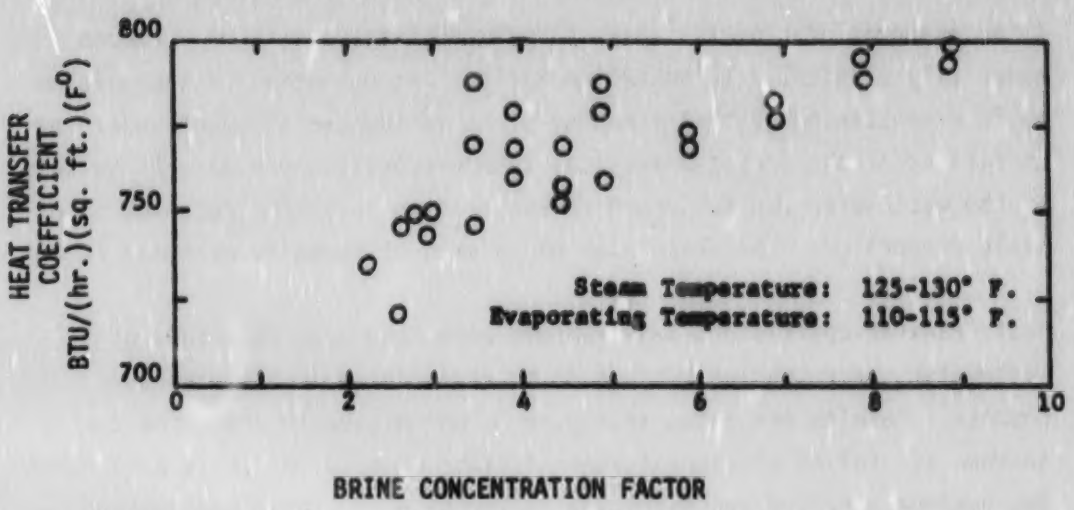
Total dissolved solids, TDS	3,500 mg/l
Total suspended solids	10 mg/l
Chemical oxygen demand, COD	250 mg/l
Oil and grease	100 mg/l
Hardness, $CaCO_3$	250 mg/l
pH	6
Calcium, Ca^{++}	50 mg/l
Magnesium, Mg^{++}	25 mg/l
Sodium, Na^{+}	300 mg/l
Potassium, K^{+}	500 mg/l
Ammonium, NH_4^{+}	200 mg/l
Iron, Fe	2 mg/l
Chloride, Cl^{-}	350 mg/l
Sulfate, $SO_4^{=}$	2,000 mg/l
Bicarbonate, HCO_3^{-}	50 mg/l

SCALE FORMATION

In an evaporation process, scale formation on heat transfer surfaces is especially critical. To determine the scaling potential of the calcium sulfate-containing synthetic rubber plant wastewater effluent under the conditions of the proposed vertical tube evaporation process, 55 gallons of the wastewater was subjected to evaporation in an Envirogenics pilot-scale evaporator. The evaporator has a product capacity of about 12 gph.

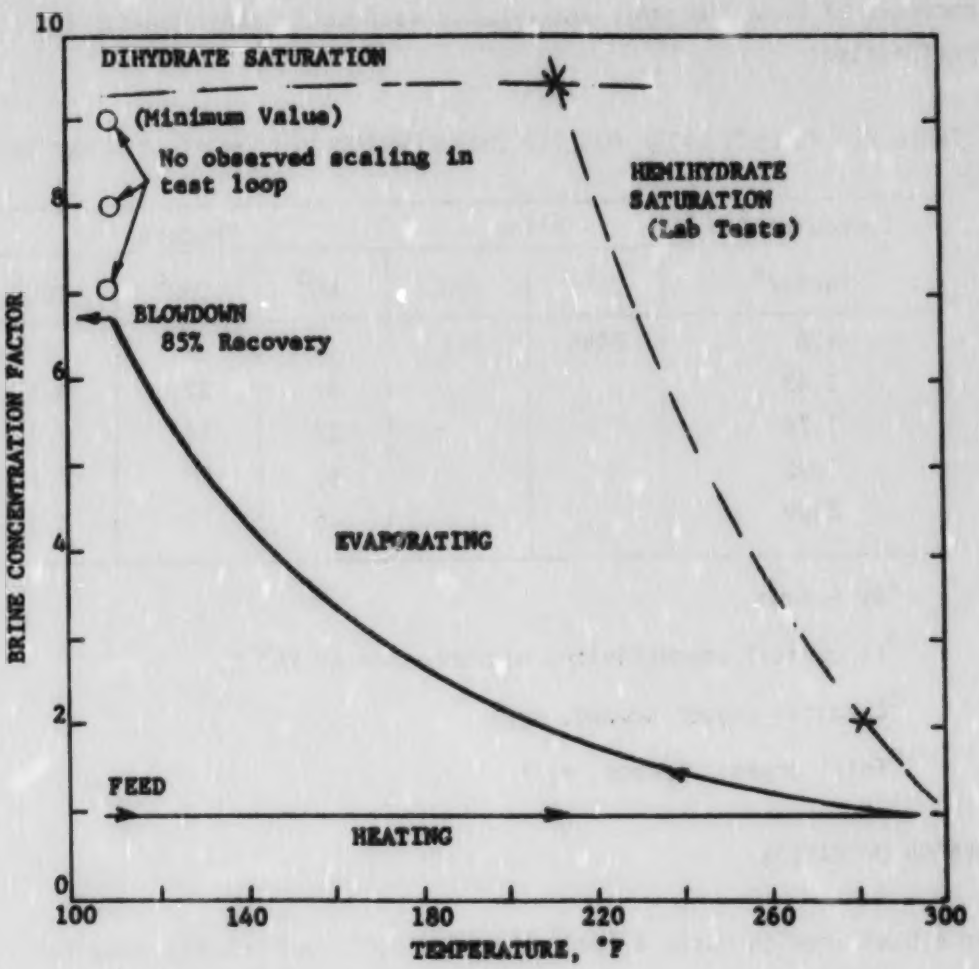
Heat transfer coefficient measurements were made over the range of wastewater concentration factors to be encountered in the proposed process. Results are shown in Figure 3, which clearly indicates the absence of scaling at concentration factors up to 9, which is well above the maximum proposed concentration factor of 6.7. These measurements were conducted at the temperature expected at the cold end of the evaporation train, namely 110^{0} F., where the wastewater is most concentrated. At this low temperature, calcium sulfate scale takes the form of the dihydrate, called gypsum. At elevated temperatures, the scale in operating plants takes on the form of the hemihydrate, or plaster of paris, which has a much lower solubility at elevated temperatures than gypsum.

Figure 3. Relationship between heat transfer coefficient and brine concentration



The concentration factors at which scale was observed to form at the higher temperatures of the process were determined in the laboratory and are presented on Figure 4. Also shown is the design process operating line which is safely within both the dihydrate and hemihydrate scaling limits. The dihydrate saturation curve was not exactly defined for the wastewater but experience indicates that it will take the form shown, i.e., relatively insensitive to temperature. In any event, saturation was not observed by the formation of scale at a brine concentration factor of 9, which is well above the expected maximum brine concentration.

Figure 4. Vertical tube evaporation operating line and scaling limits



Product water qualities were determined at selected points in the process and are presented in Table 2, together with the unconcentrated wastewater quality. The results indicate that as the concentration factor increases, and concomitantly the evaporation temperature decreases, both the inorganic and organic content of the distillate decreases. Thus the concentrations of these constituents in the aggregated product water resulting

from evaporation over the complete range of concentration factors up to 6.7 should be much lower than those indicated in Table 2, and 95-percent removals of both inorganic and organic wastewater constituents is anticipated.

Table 2. PRODUCT WATER QUALITY CHARACTERISTICS FROM PILOT PLANT DATA

Concentration Factor[a]	Brine		Product		
	EC[b]	COD[c]	EC[b]	COD[c]	TOC[d]
1.0	5495	311	-	-	-
1.43			37	32	9.1
1.76			37	34	7.4
1.92			43	28	7.1
2.69			26	18	4.5

[a]By volume

[b]Electrical conductivity, micromhos/cm at 25° C.

[c]Chemical oxygen demand, mg/l

[d]Total organic carbon, mg/l

DESIGN OPERATION

As a basic design case, a feed rate of 640,000 gpd was assumed, for which a process design was selected. Components have been sized throughout to handle 700,000 gpd feed. The plant is designed to operate without modification over a range of feed rates from 200,000 to 700,000 gpd, and recovery percentages from 40% to 85%. Table 3 summarizes important flows and other parameters for the design case.

Table 3. TYPICAL DESIGN OPERATION SUMMARY

Parameter	gpm	gpd	#/hr.
Feed rate	447	644,000	222,600
Product	307	442,000	152,000
Steam	-	-	13,900
Concentrated brine	140	202,000	70,600
Cooling water	668	963,000	333,000
Electric demand		68KW	
Water recovered		68%	

The process takes clarified wastewater at 75° F. from the holding ponds. It is heated through two stages of feedheating to a temperature of 125° F., adjusted to a pH of 4.5 with sulfuric acid, and fed to a vacuum degassifier. In the degassifier, the feed is deaerated and decarbonated, and the low-boiling organics are removed as the liquid flashes down to 123° F. The pH of the feed is brought to near-neutral when sodium hydroxide and sodium sulfite are added to scavenge trace oxygen. The feed then proceeds through the individual feedheaters where it is heated by the product water from the associated effect. After traversing the feedheater circuit, the feed enters a boiler-steam-heated exchanger, Hx301, wherein its temperature is raised to 297° F. prior to introduction to the high temperature degassifier. Upon introduction to the high temperature degassifier the hot feed flashes down to 295° F. during which the remaining volatile fraction of the oils present are removed in a vent stream.

NOTE: The venting or discharge of the volatile organics to the atmosphere may be prohibited by the new Federal, State or Local air pollution regulations. These regulations must be checked for each location in regard to the specific organic chemicals to be vented.

Since the process liquid is continuously cooling from the high temperature degassifier throughout the remaining process, further volatilization of oils will be nil. Organics appearing in the distillate during the evaporation-condensation process which follows are calculated to be less than 1 ppm in the product water if any at all are present, based on the range of oils and organics which may be expected to appear in a waste stream from the processing plant.

The two-temperature-level degassing design provides for removal of a wide range of volatile organics, not only protecting the purity of the product water, but also eliminating them from contact with the evaporation train, where some species might add significantly to corrosion problems.

The deaerated, degassified feed is introduced to the first effect evaporator upper water box where it flows through stainless steel distributors and onto the inside of the tubes in a thin film. Boiler steam on the outside of the tubes condenses giving up its latent heat to the falling film inside the tubes, causing it to boil, resulting in the generation of a quantity of steam almost equal to that condensed.

The mixture of waste liquid and product vapor issues forth from the bottom of the falling film tube bundle where the liquid falls to the bottom of the sump. The vapor is drawn out and upwards through a knitted wire mesh entrainment separator to the relatively cold outside surface of the falling film tubes of the second effect where it condenses, acting as the heating steam for the second effect. The accumulated, slightly concentrated waste liquid is withdrawn from the first effect sump and pumped to the top water box of effect two. At this point it flows through porcelain distributors and gravitates down the inside of the tubes in the same manner as for the first effect. The feed to the

second effect is superheated, relative to the second effect evaporating pressure, hence a portion vaporizes as the pressure is reduced through the distributors, contributing to production and reducing the liquid temperature to provide the potential for heat transfer.

The distillate draining from the outside of the tubes in the first effect is delivered to the product sump of the associated feedheater, along with adequate vapor to heat the feed. Each subsequent effect operates in the same manner, delivering its distillate to the feedheater product channel where it joins the accumulating stream of combined distillate (product). Since the temperature level is less in each subsequent feedheater, the accumulated distillate flashes on entering each feedheater as it is transferred in series from exchanger to exchanger through seal loops. The vaporization cools the distillate and produces vapor which condenses on the feedheater tubes, thus providing for distillate cooling and simultaneous heat recovery from the product. The combined product stream is removed from the plant at 110° F. from the sump of the coldest operating feedheater. In this manner the waste liquid is successively concentrated in each effect and finally discharged from the sump of the last effect at 110.4° F. The vapor produced in the last effect is condensed by cooling water in a direct contact barometric condenser.

Vent vapor streams are removed at appropriate points in the process, either to the atmosphere from super-atmospheric operations, or to the appropriate level of vacuum header. Venting thus continuously removes non-condensibles and avoids blanketing of the heat transfer surfaces. The vent vapors are eventually pumped to the atmosphere by steam jet-air ejectors, with cooling water-cooled barometric-type intercondensers.

The temperature profile over the plant is established and maintained by (1) the application of boiler steam to the first effect and the final feedheater; (2) the application of cooling water to the direct contact final condenser; and (3) the continuous removal of non-condensibles. It should be noted that excess surface and pumping capacity has been provided over that required for the design process case. Since the plant will have a fixed surface area, and it operates on the basis of $Q = U A \Delta T$ where both U, the overall heat transfer coefficient, and A, the heat transfer surface area, may be considered fixed, the overall temperature difference will be reduced as the rate of evaporation is reduced. Thus, the actual top temperature for the feed liquid in the design case (644,000 gpd feed and 68% recovery) is about 280^{0} F. The system is completely self-regulating as the rate of operation varies, since the first effect operating control automatically reduced the steam input as the wastewater feed flow is throttled.

An additional source of flexibility is inherent in the system. While the steam rate is automatically proportioned to feed rate, the ratio is adjustable. If a lower recovery is desired, one merely reduces the steam applied for a given feed rate (or raises the "feed to steam ratio"). The automatic control also assures that overconcentration will not occur. A pressure regulator on the steam line limits the maximum temperature on the first effect.

This particular system is also quite insensitive to variations in the quality of the feed with the provision for two-step removal of volatiles and the fact that the salt content has little effect on evaporation. This allows for performance over a wide range of waste feed conditions, which could very well occur in an industrial wastewater application such as this.

Utility requirements for the VTE are steam, electricity, cooling water, and instrument air. The plant receives 250-psig steam, dry and saturated, which is reduced to 75 psig prior to use in the first effect. The plant also receives cooling water at 75° F. and returns this cooling water, augmented by the vapor from the 14th effect, at 106° F. Three-phase electric power at 60 cycles and 480 volts is required for all pump drivers and/or to energize a step-down transformer to supply single phase power for lighting, control, and fractional HP motor drives.

CONSTRUCTION

Field construction was begun during June 1971, and the installation completed during February 1972.

Figures 5 through 7 are photographs showing various stages of construction. Figure 5 shows the construction site as of July 30, 1971, with foundations completed and effect circulating pumps set in place. In Figure 6, taken November 5, 1971, the evaporator vessel has been set with the feedheaters placed underneath; the vacuum degassifier, control building, and cooling tower are seen in the background. Figure 7 shows the completed facility.

The mechanical contractor for the project was Ref-Chem Corporation, Odessa, Texas.

Total capital cost for the VTE installation was $1,184,783. See section entitled Economics for a breakdown of the capital costs including the cost of the solar evaporation ponds and pilot plant equipment.

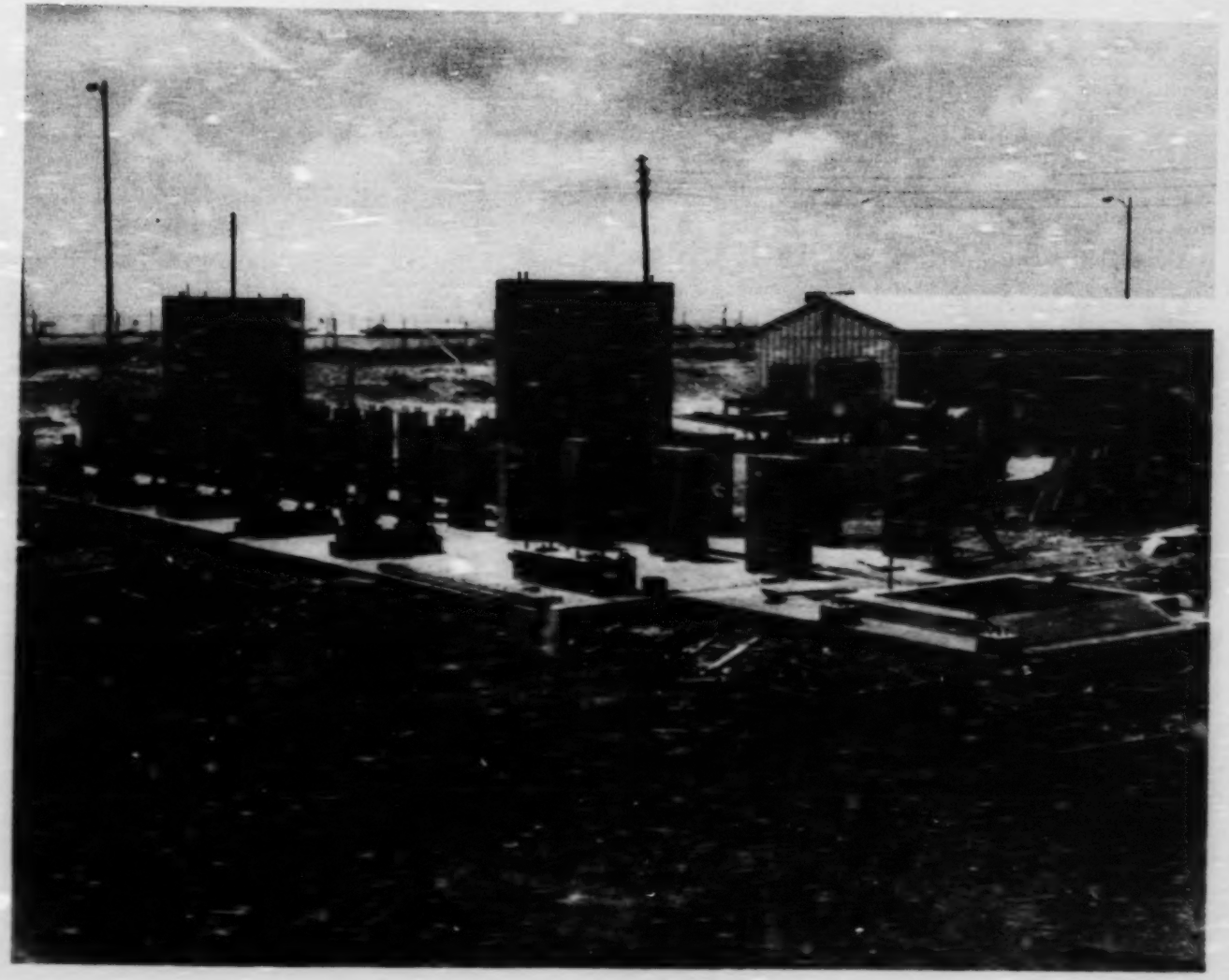


Figure 5. Construction site, July 30, 1971

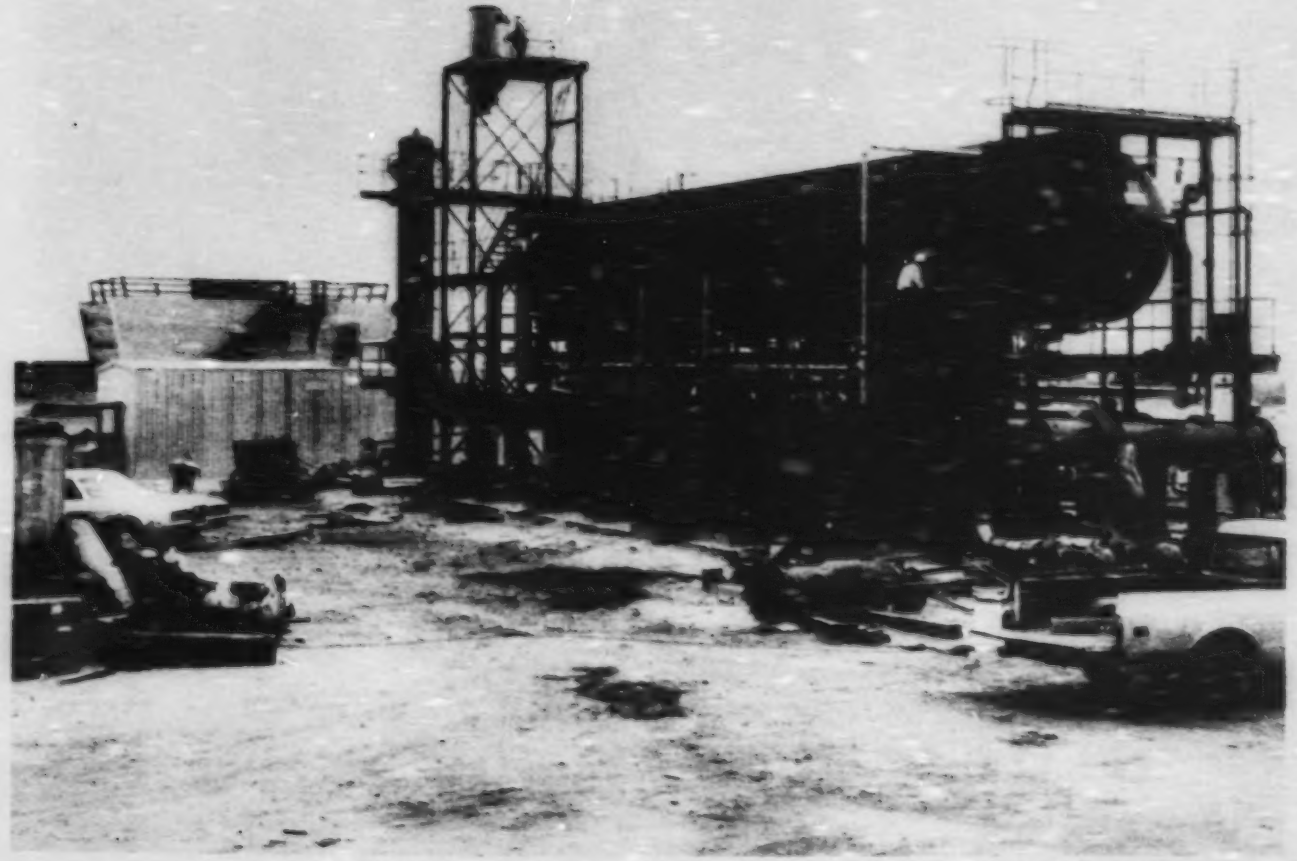


Figure 6. Construction site, November 5, 1971

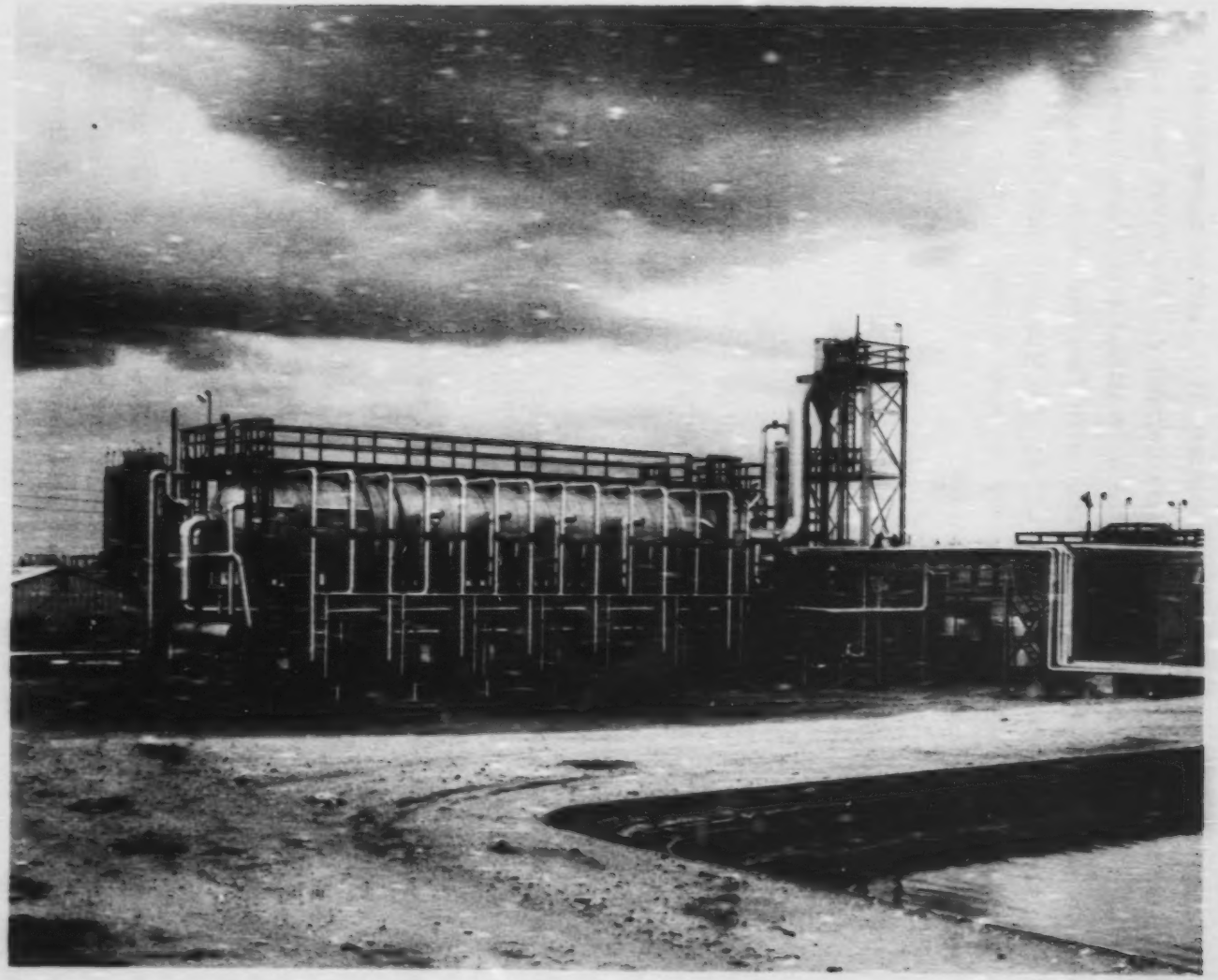


Figure 7. Completed facility

OPERATION

INITIAL OPERATION

The VTE was brought on stream during January 1972, and normal mechanical and instrumentation run-in problems were corrected. The first heat and material balances calculated around the unit showed that thermal performance exceeded design specifications, with feed and steam rates at design, the product rate was higher as was the resultant gain ratio. Averaged data for a 70-hour performance run made during the first week of February 1972 is presented in Table 4.

Table 4. AVERAGED DATA FEBRUARY 1972 PERFORMANCE RUN OF 70-HOUR DURATION

Parameter	Measured value	Design value
Feed rate	623,500 gpd (432 gpm)	644,000 gpd (447 gpm)
Production	475,488 gpd (330 gpm)	442,382 gpd (307 gpm)
Steam rate	13,976 lb./hr.	13,900 lb./hr.
Gain ratio $\frac{\text{lbs. of water}}{\text{lbs. of steam}}$	11.7	10.6
Feed concentration $\frac{\text{feed}}{\text{brine}}$	4.2	3.19
Water recovery	76%	68%

The eight-day average for period June 27 to July 7, 1972:

Feed rate	300.5 gpm
Production rate	185.4 gpm
Steam rate	170 pounds per min. = 10,200 pounds per hour
Gain ratio	10.2
Water recovery	62%

During the period of February to June 27, 1972, operation of the unit was sporadic as we attempted to solve some of the operational and fouling problems. Table 5 shows the product water analysis for four of these short runs.

Table 5. PRODUCT WATER COMPOSITION DURING INITIAL OPERATION

	Product Water Composition			
	2/6/72	3/26/72	3/27/72	3/29/72
pH at 22° C.	9.2	6.0	7.2	7.0
Specific conductance, micromhos/cm	55.5	60.4	53.4	32.0
Chloride, Cl ppm	2.8	1.1	1.2	1.2
Sulfate, SO_4 ppm	ND	9.9	25.0	6.0
Phenol ppm	0.88	1.12	0.45	0.42
TOC ppm	10.3	13.0	51.5	12.5
Ammonia nitrogen, as N ppm	10.0	0	10.0	6.8

AMMONIA IN PRODUCT

The biggest problem with the quality of the product water during this period was the carryover of the ammonia. Ammonia is detrimental to the synthetic rubber polymerization reaction. The cause of this was traced to the inability of the degassification system to reduce ammonia in the feed to design levels. Experimental work led to the addition of an acidification step after the high temperature degassifier to maintain the feed pH at less than 7.0, thus maintaining the ammonia in solution as ammonium ion. The problem was further reduced by a changeover from ammonium hydroxide to sodium hydroxide neutralization in a waste treatment step upstream of the VTE and ammonia level in the wastewater feed dropped to about 50 ppm. These changes resulted in production of water of such quality as to allow recycling to the synthetic rubber process, with ammonia held well under 10 ppm.

Operation of the unit after June 27, 1972, utilized the above operational changes; and generally the ammonia was approximately 5 ppm or below, and the total dissolved solids were below 10 ppm. This data is included in Table 6 along with the operating data.

FEEDHEATER FOULING/CORROSION

During March 1972, the first indication of feedheater fouling was evidenced by a decay in VTE efficiency as measured by the gain ratio. This necessitated the first of what became an almost monthly routine of shutting down the VTE unit, disassembling the feedheaters, scouring the tubes with a high-pressure (10,000-pound-per-square-inch) water jet, and reassembling of the feedheaters. Each such cleaning was quite satisfactory in producing a clean metal surface, but as the 70-30 copper-nickel tube surface progressively roughened from corrosive attack, the technique became less effective. The spiral indentations precluded drilling of the tubes, but the deposits being soft were readily removed and did not require such severe treatment. Figures 8 through 14 are microphotographs of corroded feedheater tubes.

By July 1972, a randomly selected feedheater tube from the hot end of the VTE showed corrosion penetration of one-half the tube wall. By September 1972, some corrosive attack of the vertical tubes was noted; although, at an apparently lower rate than the attack on the feedheater tubes. Throughout the program, the type 316 stainless steel piping, pumps, and cladding appeared unaffected by corrosion.

Analysis of samples taken from the feedheater tubes are shown in Tables 6 thru 10. The corrosion of the copper was undoubtedly caused by the presence of the sulfide ion with the major foulant being copper sulfide and later on organic materials. The source of the sulfide ion was traced to organic sulfides used in the manufacturing process and anaerobic bacterial action in the effluent feed holding pond reducing the sulfates to sulfides.

The total concentration of these sulfide ions ranged from 2 to 5 ppm, with some much higher and lower levels at times. This amounts to 0.44 to 1.11 pounds per hour or 10.7 to 26.7 pounds per day in the 644,000 gallons per day of effluent feed.

Table 6. ANALYSES OF SAMPLE TAKEN FROM PREHEATER TUBES DURING MAY 1972

Silica, SiO_2	0.03%
Copper, as Cu	58.3
Nickel, as Ni	4.1
Iron, as Fe	0.14
Chromium, as Cr	0.31
Sulfur, as S	12.7
Organics	Not over 20

Table 7. ANALYSES OF SAMPLES TAKEN FROM PREHEATER TUBES IN JULY 1972

Sample Source	HX301	1st Effect Tube Sheet	HX315
Copper, as Cu	55.8%	41.6%	43.7%
Nickel, as Ni	1.1	0.3	18.4
Sulfur, as S (total)	15.9	12.3	4.6

Table 8. ANALYSES OF SAMPLES TAKEN FROM PREHEATER TUBES IN SEPTEMBER 1972

Sample	Light, Granular Material	Moist, Dark Appearing Material
Organic matter	44%	38%
Acid insolubles	1.9	0.8
Sulfur, S	5.7	3.9
Nickel, Ni	0.7	0.5
Aluminum, Al	2.5	3.9
Copper, Cu	0.7	0.3
Iron, Fe	0.6	0.3

Table 9. ANALYSES OF SAMPLES TAKEN FROM PREHEATER TUBES IN SEPTEMBER AND OCTOBER 1972

(During a period in which chlorination appeared to be oxidizing sulfides but aggravating fouling by creating organic compounds more likely to deposit on the tube surfaces.)

Constituent	Weight Percent		
	11 Sept. Samples		23 Oct. Sample
	No. 1	No. 2	No. 3
Loss on Ignition	44.0	38.0	21.6
Aluminum (as Al_2O_3)	4.7	7.4	59.1
Calcium (as Ca)	12.6	4.6	0.03
Magnesium (as Mg)	2.1	2.5	0.32
Carbonate (as Co_3)	NR	NR	4.6
Acid Insolubles (SiO_2)	1.9	0.8	2.6
Copper (as Cu)	0.7	0.3	NR

Continued on next page

Table 9 (continued). ANALYSES OF SAMPLES TAKEN FROM PREHEATER TUBES IN SEPTEMBER AND OCTOBER 1972

Constituent	Weight Percent		
	11 Sept. Samples		23 Oct. Sample
	No. 1	No. 2	No. 3
Nickel (as Ni)	0.7	0.5	NR
Iron (as Fe)	0.6	0.3	NR
Phosphate (as PO_4)	NR	NR	0.05
Sulfur			
Total (as SO_4)	17.1	11.7	4.4
Sulfide	NR	NR	none detected
Elemental (as S)	NR	NR	<0.003
Material extractable in MIBK/ether (1/1)	NR	NR	1.2
Total Carbon	NR	NR	3.55
Organic Carbon (total carbon less carbonate carbon)	NR	NR	2.62
Total Hydrogen	NR	NR	2.42
Total Nitrogen	NR	NR	0.75

Table 10. ANALYSES OF SAMPLES TAKEN FROM CORRODED EVAPORATOR TUBES DURING MARCH 1973

Emission Spectrographic Data
(approximate wt %)

Sample Color	Tube 1			Tube 2		
	White	Blue	Black	White	Blue	Black
Si	0.05	0.05	0.1	0.05	0.1	0.1
Mn	0.1	0.1	0.1	0.05	0.5	0.1
Cr	0.1	0.05	0.1	0.1	0.05	0.05
Mg	0.5	0.5	0.5	0.5	0.5	0.5
Fe	0.8	0.8	0.8	0.8	0.8	0.8
Ca	0.1	0.5	0.5	0.5	0.75	0.5
Ag	0.05	0.05	0.1	0.05	0.05	0.05
Ni	1.0	1.0	1.0	1.0	1.0	1.0
Al	5.0	3.0	3.0	5.0	3.0	3.0
CU	Major component greater than 90%					

Atomic Absorption

Ni	1.14%
Cu	39.6
Ca	1.7

Figures 8 thru 14 are microphotographs of corroded preheater tubes.

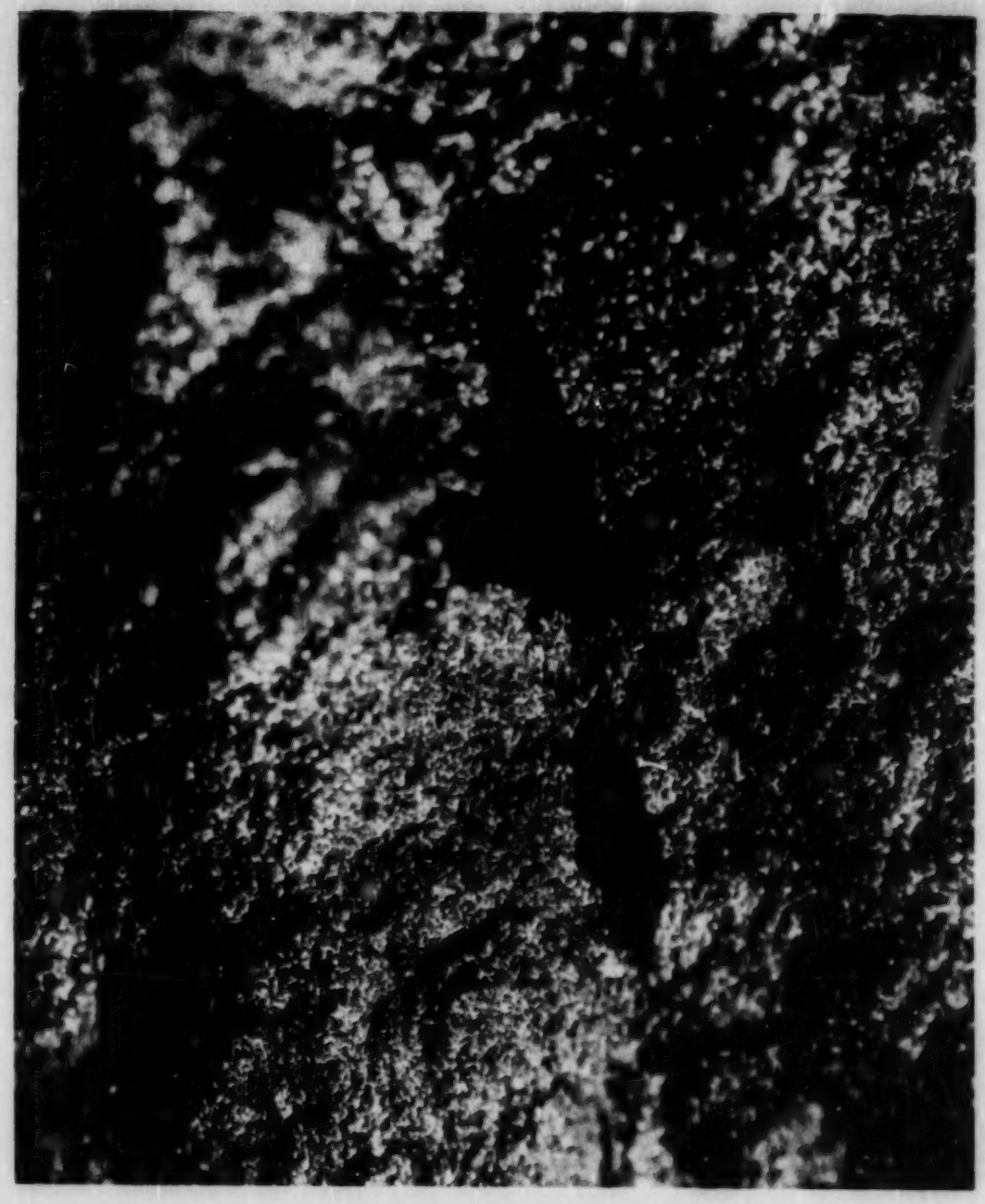


Figure 8. Holes in tube

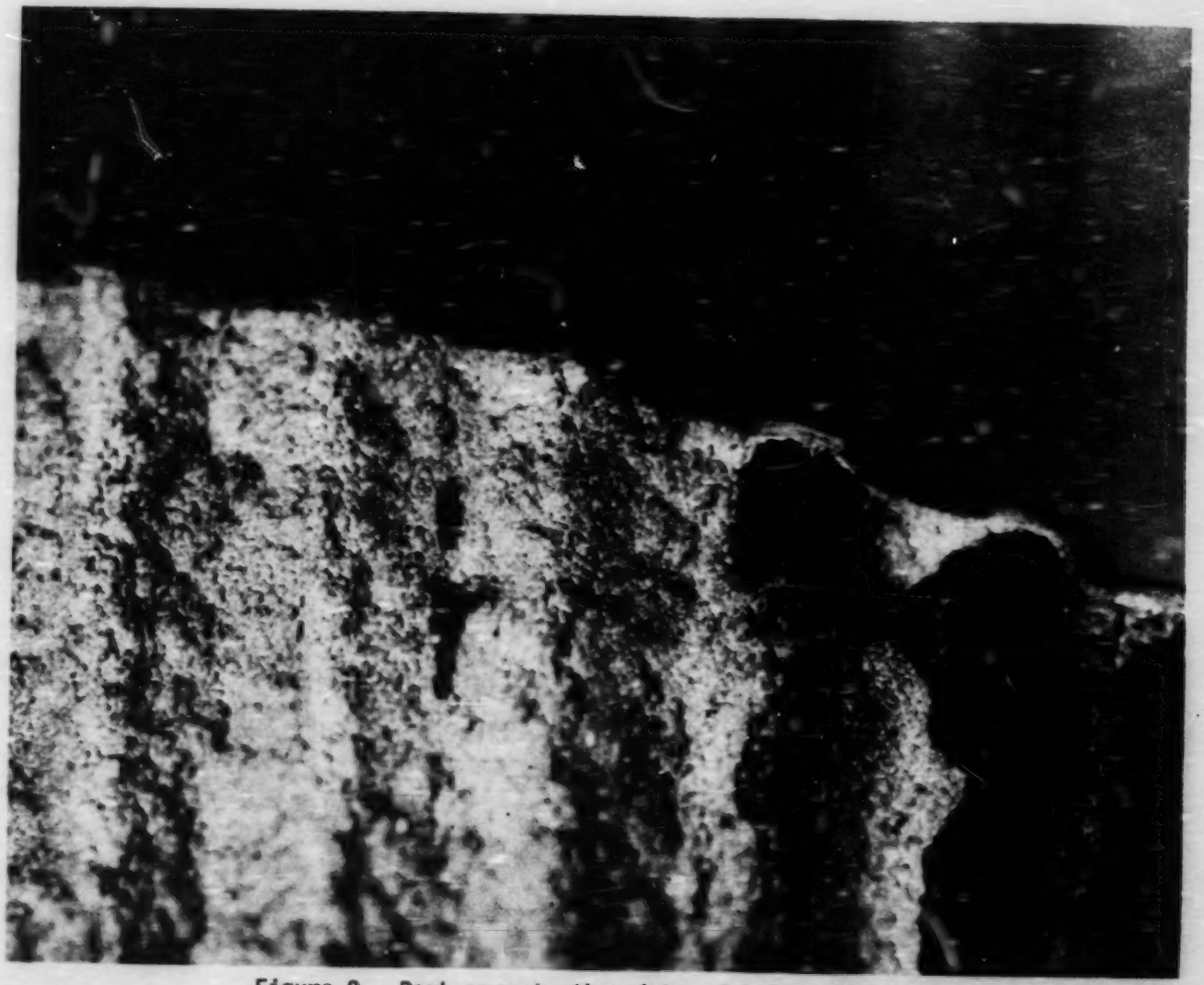


Figure 9. Dark area to the right of hole. Dark area was material not removed by NaCN solution

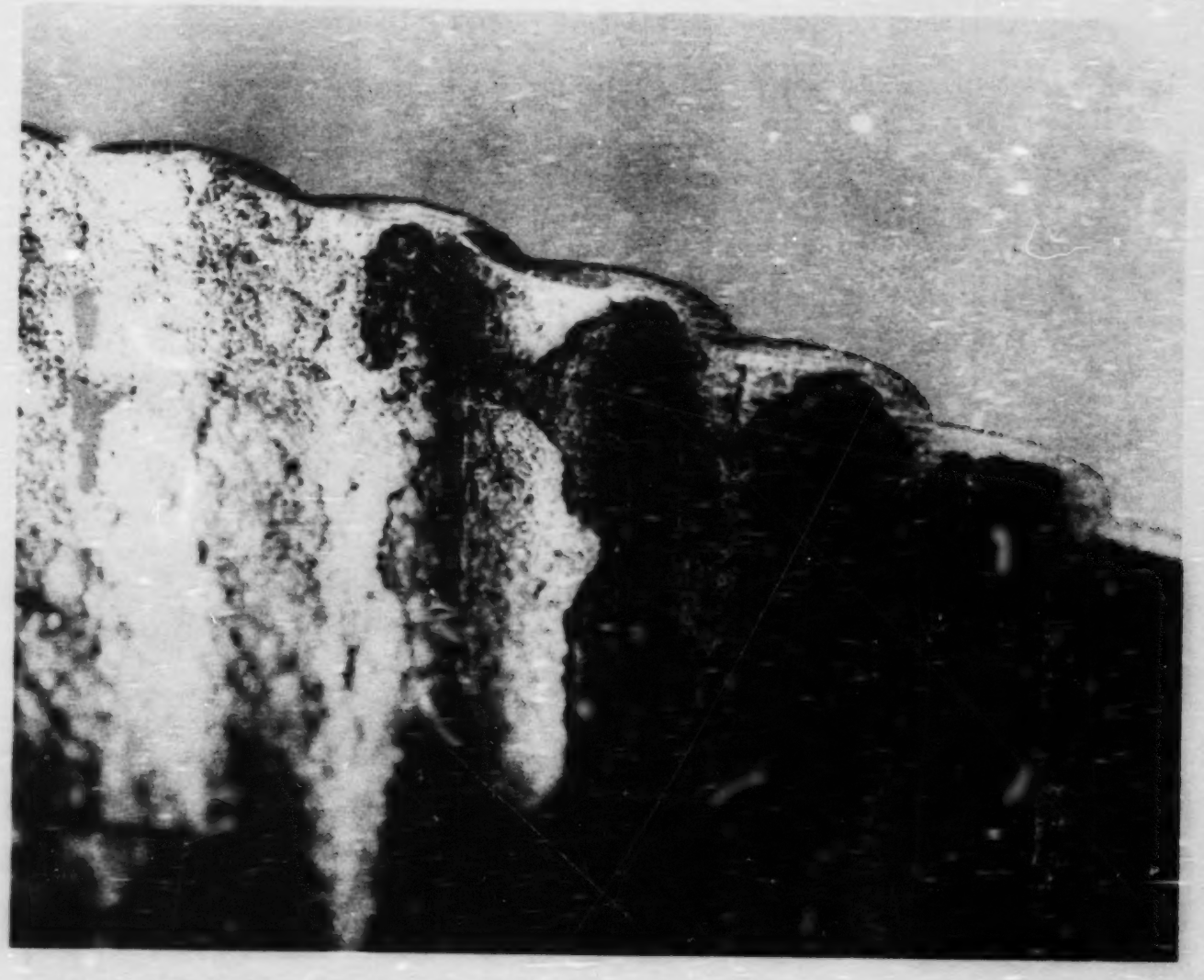


Figure 10. Same section as Figure 9 but showing more of the dark area

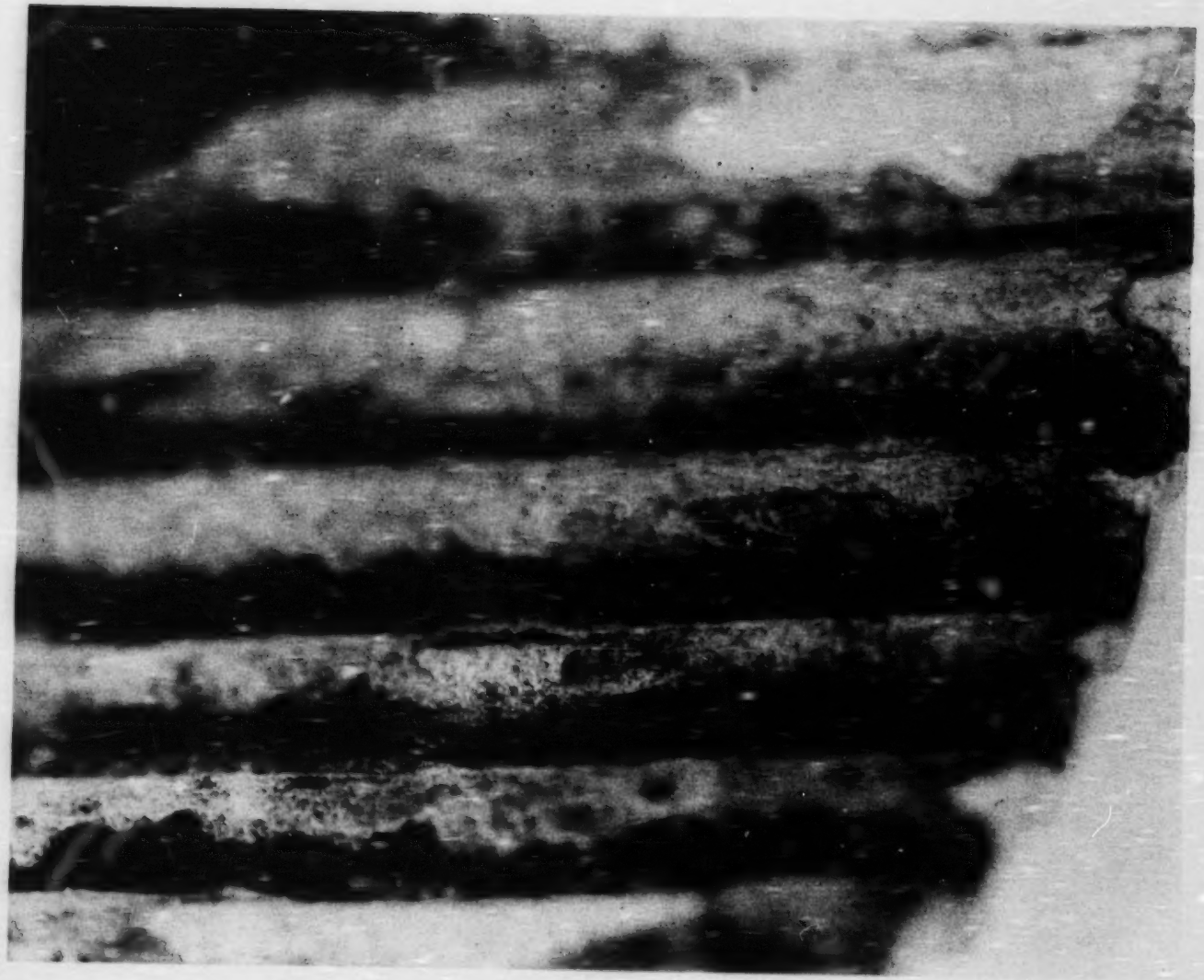


Figure 11. General view showing surface condition

Figure 12. End view of a cut tube

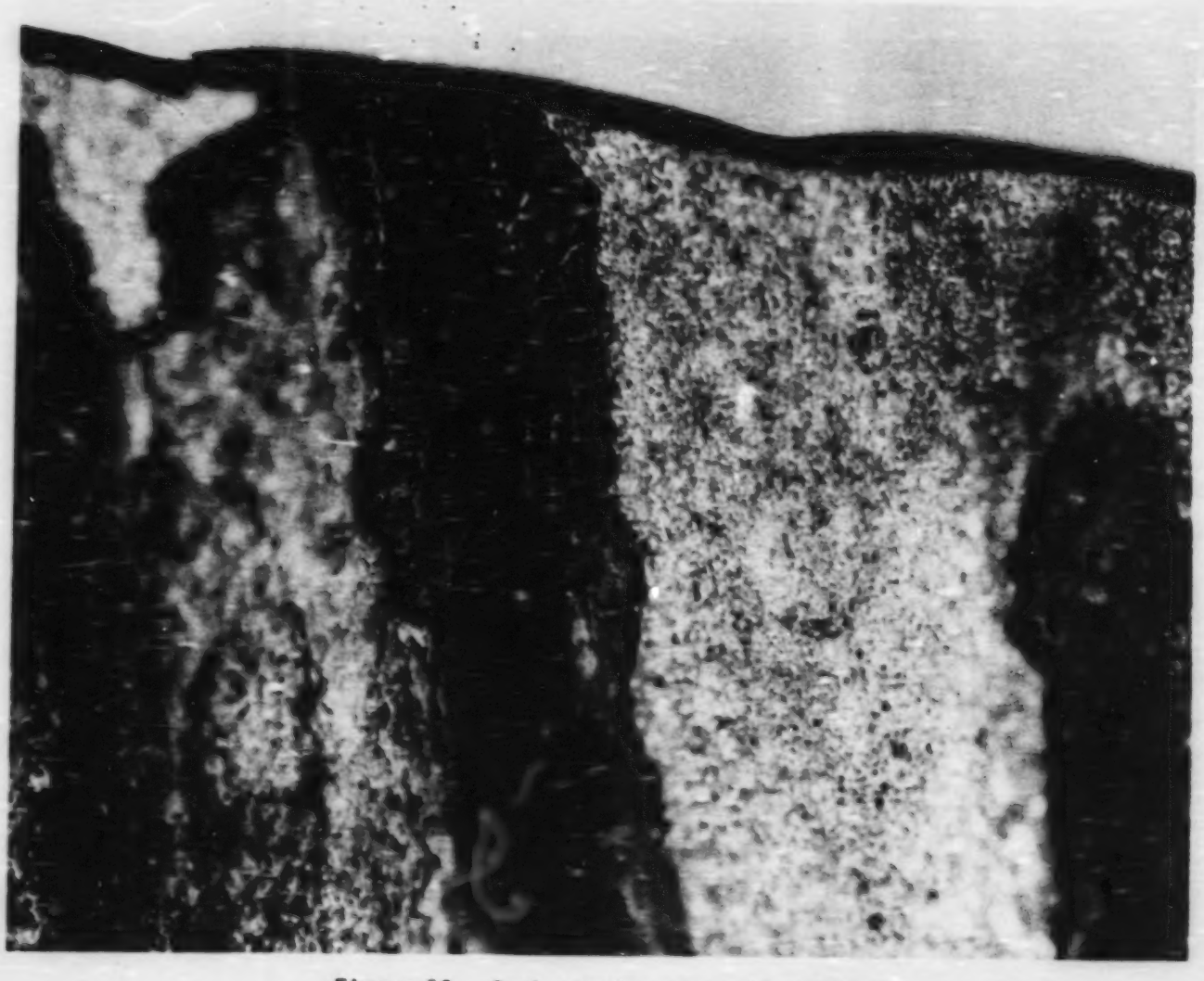


Figure 13. Dark section corrosion product

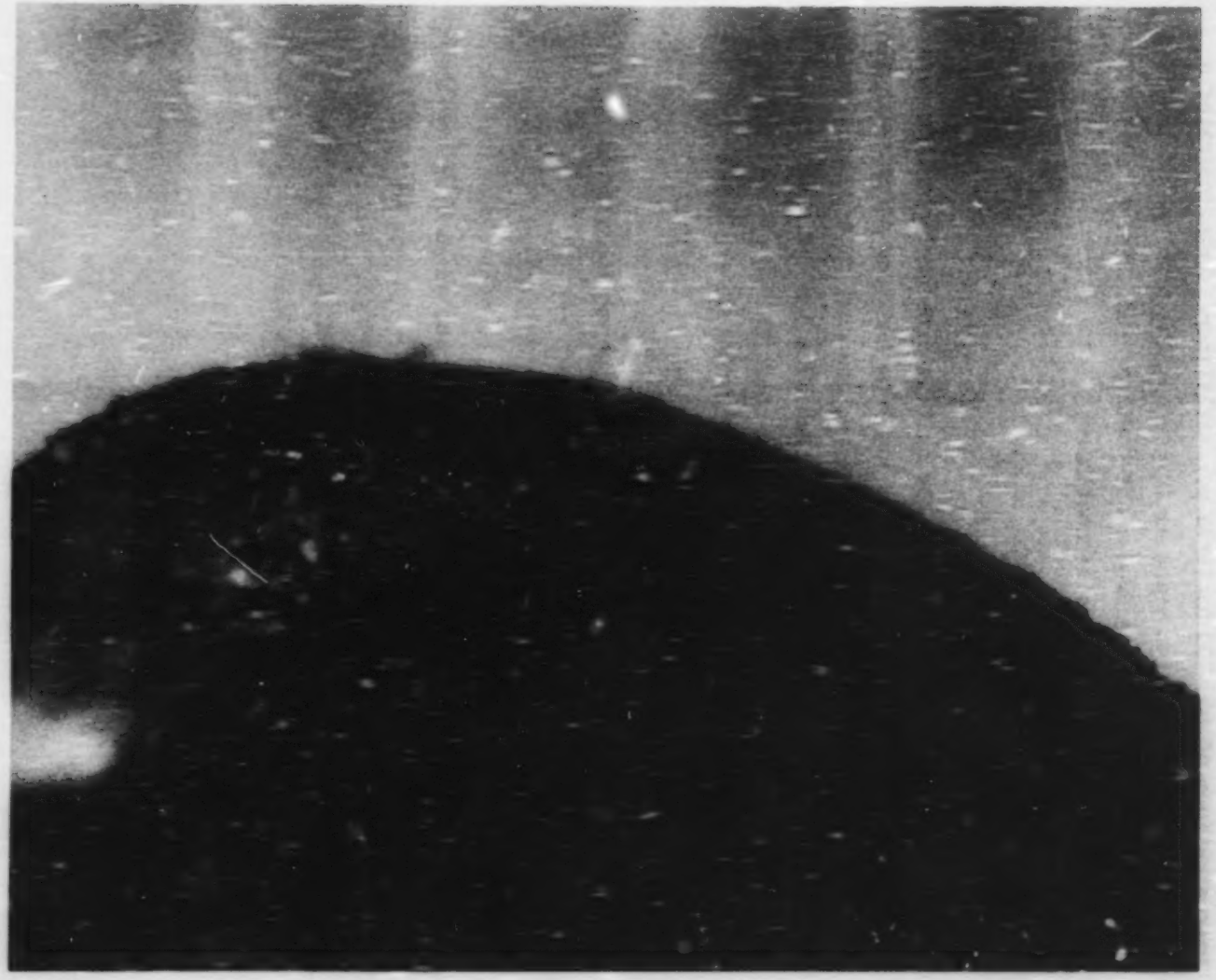


Figure 14. End view

BIOCIDE TREATMENT

Sulfate-reducing micro-organisms were identified in the feed stream in concentrations of 500-700 bacteria per milliliter. This indicated a bacteriological mechanism might be triggering the problem and the feed was treated with a biocide over a one-month period. The results were negative and the biocide was discontinued.

WASTEWATER FEED CHLORINATION

Some early potential solutions to the sulfide problem, including such techniques as stripping hydrogen sulfide from the feed stream with air, feed pH adjustments, slow sand filtration, charcoal filtration, aerating the holding ponds, flocculation, time stabilization, and reaction with iron or copper or zinc salts, yielded to chlorination of the feed stream as being the most efficient and expedient technique available to eliminate sulfide in the feed. The theory behind chlorination of the feed is that sulfide constituents are oxidized to innocuous sulfates which do not contribute to corrosion of the copper alloy tubes.

A chlorinator was installed on the wastewater feed stream in early July 1972 and was operated during most of the balance of the demonstration period. Contact times were varied between 30 seconds and 30 minutes, while the chlorine residual was held between 0.4 to 0.8 ppm. Sulfide levels of generally less than 0.2 ppm were attained. The remaining chlorine was eliminated by reaction with the sodium sulfite normally added to the feed after deaeration as an oxygen scavenger, thus reducing

the possibility of chlorine corrosion of the tubes. First indications were that chlorination substantially reduced both fouling and corrosion; however, qualitative evaluations of fouling rate were clouded somewhat by frequent periods of operation at low rates, which would tend to increase fouling. Early optimism as to chlorination's efficacy in reducing fouling waned, however, as the experiment continued, until it was ultimately believed that chlorination was contributing to the fouling. The mechanism for this enhancement of fouling is believed to be that some organics were being chlorinated and the reaction product was depositing on the tubes. Feed chlorination was continued in spite of the increased fouling, however, as it was fairly certainly established that chlorination did lower the rate of corrosion, although not eliminating it.

Reliance on chlorination exclusively, even if efficacious, was not considered the ultimate solution to the fouling/corrosion problem of the copper alloy tubes. In addition to a considerable corrosion potential in feeding large amounts of chlorine, chlorination does not oxidize elemental sulfur, organic sulfides, or some organic compounds which may be involved in the mechanism of fouling and which, although unaffected by chlorination at ambient temperatures, break down at the higher, evaporator temperatures. Additionally, there was a possibility that chlorination might produce elemental sulfur which could react with the copper.

BIO-OXIDATION OF THE WASTEWATER FEED

Bench-scale studies indicated bio-oxidation was a feasible means of oxidizing sulfides as well as oxidizing the organics in the feed. Bio-oxidation does not suffer from the drawbacks attending high-level chlorination and will oxidize sulfide-producing compounds that may be unaffected by chlorine but which break down at the high temperatures in

the VTE. Additionally, organic constituents would likewise be oxidized. Based on these studies, a 30-gallon-per-minute, rock-fill, pilot-scale trickling filter was installed at the VTE site to treat a sidestream of the VTE feed.

In the trickling filter pilot plant operation, wastewater is taken from the existing holding pond and pumped to the top of the trickling filter to flow over the rock fill, contacting the biological growth on the rock surfaces, with oxidation of sulfide and organic compounds. Air is pulled through the filter cocurrently with the wastewater. From the trickling filter, the treated wastewater flows by gravity to a settling tank for removal of any of the biological growth that has sloughed from the trickling filter rock fill. From the settling tank, the wastewater flows to a feed tank from which it is pumped through a sand/gravel filter which simulates the existing traveling media filter in the plant system. After filtration, the treated feed is pumped through a series of six, single-tube heat exchangers which simulate the range of conditions experienced in the VTE feedheaters. This arrangement of heat exchangers is tubed with both 70-30 copper-nickel tubes and titanium tubes and instrumented so that the rate of fouling can be determined.

It was contemplated that bio-oxidation, if successful in the pilot scale and installed on a full scale would be followed by low-level chlorination to scavenge residual sulfides.

The trickling filter was put on stream in December 1972 after inoculation with domestic sewage. After several weeks of operation, it became apparent that no greater than 31 percent reduction in COD could be achieved (Table 11). More recent and more extensive bench-scale experiments confirmed that bio-oxidation alone is apparently inapplicable to the particular system of organics in the wastewater; although, sulfide reductions of over 90 percent have been realized in bench scale tests.

The latter finding lost significance, due to the decision to retube the VTE with titanium tubes (see Page 47). The pilot plant was kept operating during the shutdown of the VTE for retubing and studies with it alone or in combination with one or more of the other feed pretreatment techniques was continued. The test bank of heat exchangers was utilized in subsequent pretreatment evaluations to determine rate of fouling and corrosion and also for tests on methods of cleaning.

Table 11. TRICKLING FILTER COD REDUCTION

Date	Feed COD	Outlet COD	% Reduction
12/7/72	472	407	15.9
12/18/72	323	252	22.0
12/22/72	372	272	26.9
12/27/72	391	282	27.9
1/3/73	338	268	20.7
1/5/73	378	258	31.8

ON-STREAM FEEDHEATER CLEANING

A parallel thrust in the program to solve the feedheater problems was a search for a suitable method to clean the feedheaters without the necessity of shutdown or disassembly, or, at least, no more than a brief interruption in operation. One approach investigated was an on-stream mechanical cleaning system. This system automatically and continuously circulates abrasive sponge rubber balls through the heat exchanger tubes during operation to wipe accumulated matter from the walls. This technique is commonly used on the condensers in West Coast power plants and one such system is being used in a seawater desalination plant located at Rosarito Beach, Tijuana, Mexico. A manual system was installed and operated briefly. Although it did appear to have some beneficial value, continuous use was prevented by lack of mechanical facilities to handle

the balls and the unsuitability of the sponge rubber balls for operation at the highest temperatures in the evaporator. Both of these problems were studied, but the cost involved compared to the anticipated results did not seem to warrant the expenditure.

CHEMICAL CLEANING

In addition to the mechanical cleaning method described in the previous paragraph and the high-pressure water flushing method, various chemical cleaning methods were investigated with the hope that the cost of dismantling the unit could be avoided. The most promising of these were:

1. A 1% inhibited hydrochloric acid
2. Sodium hypochlorite

The inhibited hydrochloric acid gave a spectacular improvement in the gain ratio the first time it was used. Two subsequent attempts to acid clean the unit did not result in any increase in thermal efficiency. Due to the possible damage from corrosion to the unit, acid treating was discontinued. The sodium hypochlorite proved highly successful in the laboratory in removing feedheater tube deposits; however, when it was used on the feedheater in the plant, it was only slightly effective. Other chemical methods considered involved cyanides; but as the resultant mixtures were highly poisonous and would create a very difficult disposal problem, they were not tried in the plant.

SUBSTITUTE MATERIAL OF CONSTRUCTION

The original VTE feedheater tubes were specified as 70-30 copper-nickel which is highly resistant at the temperatures encountered in operation to the 10 components originally found by analysis in the Odessa wastewater (ref. Table 1). This choice was based, in part, on short-term

pilot-scale runs on a quantity of actual plant wastewater. Although the reason is immaterial at this point, it is apparent that either the original waste analysis was not complete enough to detect 2 to 5 ppm of sulfide, the wastewater characteristics changed, the corrodant/foulant was lost through reaction or evaporation when drums of waste were transported to the pilot plant, or some combination of factors. As noted, further work has pointed to small quantities (2 to 5 ppm) of sulfide compounds as the major factor in the corrosion/fouling problem.

In December 1972, it was discovered that the tubes in all but the two feedheaters at the coldest end of the VTE had corroded to such an extent that replacement was required. These tubes were replaced early in 1973 with titanium tubes with the same spirally indented shape as the original copper-nickel tubes. Titanium is known to have excellent corrosion resistance to all presently identified wastewater components including sulfides. Stainless steel also would not be subject to sulfide attack; however, it would not be particularly satisfactory as a tube material because of its high susceptibility to stress corrosion cracking in a chloride environment.

With the expected elimination of feedheater tube corrosion, it was assumed the rate of fouling would be reduced, since the contribution of copper sulfide would be eliminated. Fouling was not expected to be eliminated entirely. Test tubes of titanium that had been installed in some preheaters during the last few months of operation still showed some evidence of fouling when inspected. Efforts were continued, consequently, to find a feasible method of eliminating foulants through pretreatment or through reducing the effect of the fouling by on-stream cleaning.

During the feedheater retubing operation, the vertical evaporator tubes were inspected and found to be also in a severely corroded state. Plans were made to also retube these with titanium tubes before the VTE was again placed in service.

OTHER PRETREATMENT TECHNIQUES

In an effort to improve the bio-degradability of the wastewater constituents, a laboratory investigation was carried out on ozonation of the feed to determine if such treatment would enhance the bio-degradability of some of the organic molecules. Preliminary results indicated BOD and COD reductions of 75 percent, with concurrent improvements in color and turbidity. These results when compared to simple bio-treatment suggested that it might be feasible simply to ozonate the feed and produce a feed stream acceptable to the VTE without any biological treatment.[2]

An ozonation pilot operation was constructed using the test bank of six single tube heat exchangers from the bio-oxidation tests. A 1.7% mixture of ozone in oxygen was generated at the plant site, and the feed stock was treated at a rate of 200 ppm of ozone. While this treatment did reduce the COD, fouling continued with the foulants now consisting of iron oxide and aluminum salts. Fouling was much greater when the plant was making alum (aluminum sulfate) coagulated rubber. Several chemical after-treatments were tried to eliminate these new foulants, namely, the addition of a dispersant, sodium sulfite, and sodium carbonate. The net result of the ozonation tests was that ozonation by itself would not eliminate the fouling.[3]

An activated carbon adsorption pilot run was made. The carbon also reduced the COD as did the ozonation; however, the iron and aluminum foulants still remained. The aluminum source is the alum (aluminum

sulfate) used to coagulate the synthetic latex and also used throughout the plant to neutralize alkaline wastes and latex spills. The iron was traced to the acid used in the pretreatment of the wastewater and to complex iron organic compounds in the wastewater which were broken down by both the ozonation and carbon adsorption and also by bacteriological action. It is highly probable that these foulants had always been present but were masked by the organic foulants and copper sulfide.[4,5]

CALCIUM SULFATE SCALE

One important observation over the period of operation of the VTE is that essentially no calcium sulfate scale of either the dihydrate or hemihydrate type formed on the evaporator tubes, in spite of wide variations in and unsettled conditions of operation. This confirmed calculations and studies made during the design of the evaporator.

OPERATING DATA

Table 16, Pages 65 to 89 in the appendix, presents the operating data and chemical analyses for the period from June 27, 1972, through July 25, 1974.

After the unit was restarted on July 6, 1973, after the retubing operation, the laboratory analyses were discontinued as the reliability of the process control instruments had been proven and could be depended upon to sound an alarm and bypass the product receiving tanks if the product water exceeded 10 ppm total dissolved solids.

It was anticipated at project inception that rather complete and extensive data could be gathered during the project regarding operating and maintenance costs; however, the continuing problems with heat exchange surface corrosion and fouling precluded this.

With the exception of several short periods immediately following tube cleaning, the efficiency of the VTE, as indicated by the gain ratio, was below 10.0 due to feedheater fouling. The quality of the product water was very high, and the unit constantly produced water under automatic control of less than 10 parts per million of dissolved solids. The product water was used in the manufacture of synthetic rubber throughout the operating period.

Data in Table 16 are based on the following:

1. Liquid flows are the average based on meter readings taken over a 10- to 20-minute period and are in gallons per minute. Steam flows likewise are the average flow over the same 10- to 20-minute period and are in pounds per minute.

2. The product total and the gain ratio include the steam condensed in the first effect. This is the proper method of accounting, since an almost equal quantity of steam is taken from the final effect and is condensed in the barometric condenser, thus being sent to the cooling tower as make-up water. (An almost equal quantity is evaporated in the cooling tower.) Thus, the quantity of steam used is not counted as product. If a surface condenser instead of a barometric condenser had been utilized, the steam from the last effect would be collected with the product, and the proper method of accounting would be to deduct from the product total the quantity of steam fed to the first effect.

3. The analytical data shown is for grab samples taken at the same time the operating data was taken, approximately 9:00 a.m. on the dates indicated.

Figure 15 shows the decay in the gain ratio which is directly related to the degree of fouling for the periods of continuous operation for the entire operating period. Gain ratios of 10.0 or better are needed to keep the operating costs reasonable. At a gain ratio of 10, 1,000 pounds of steam will produce 10,000 pounds of water or 1,205 gallons making the fuel cost for the water per 1,000 gallons 83% of the cost of steam per 1,000 pounds.

Figure 15. Gain Ratio vs. Time

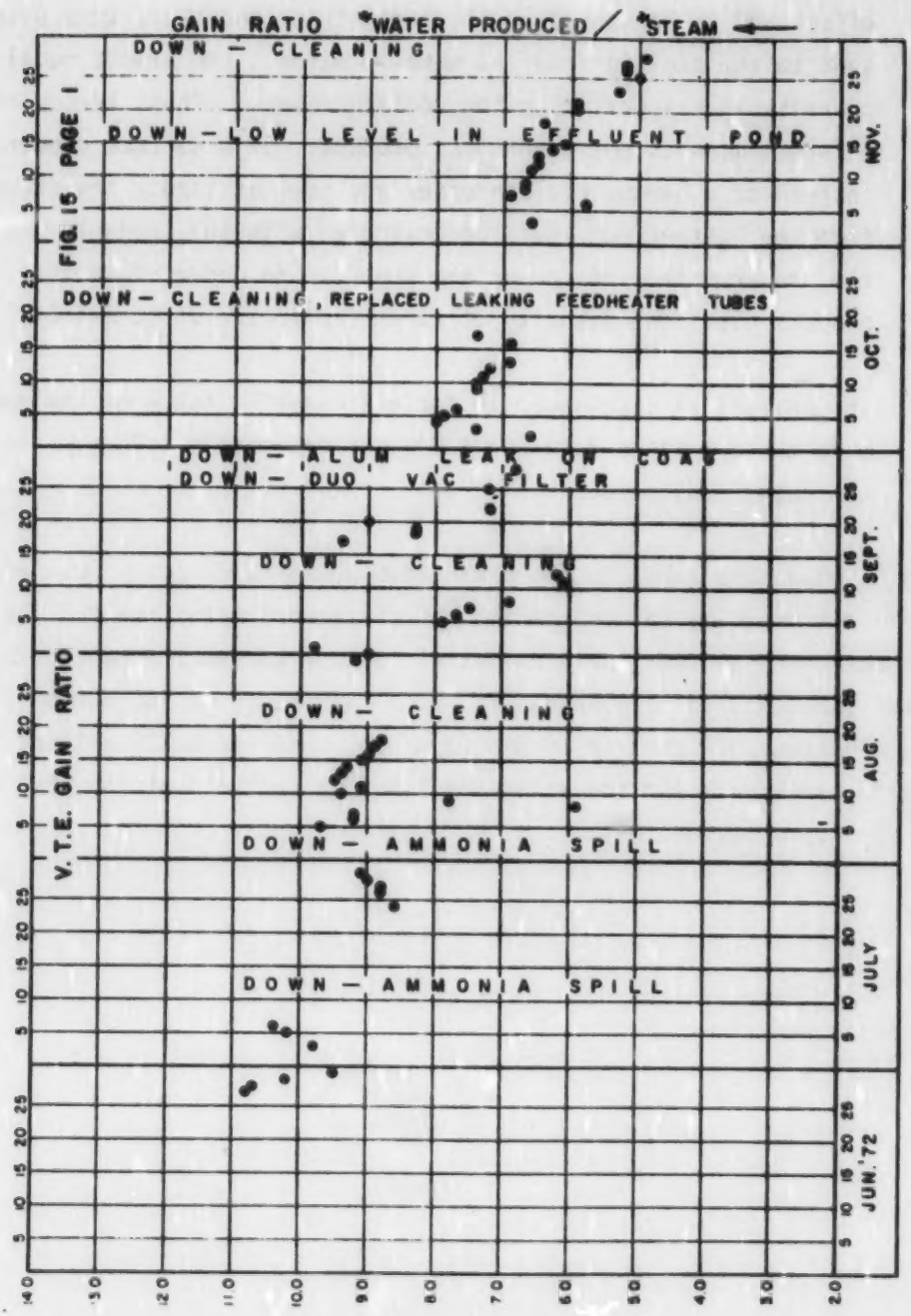


Figure 15 (continued). Gain Ratio vs. Time

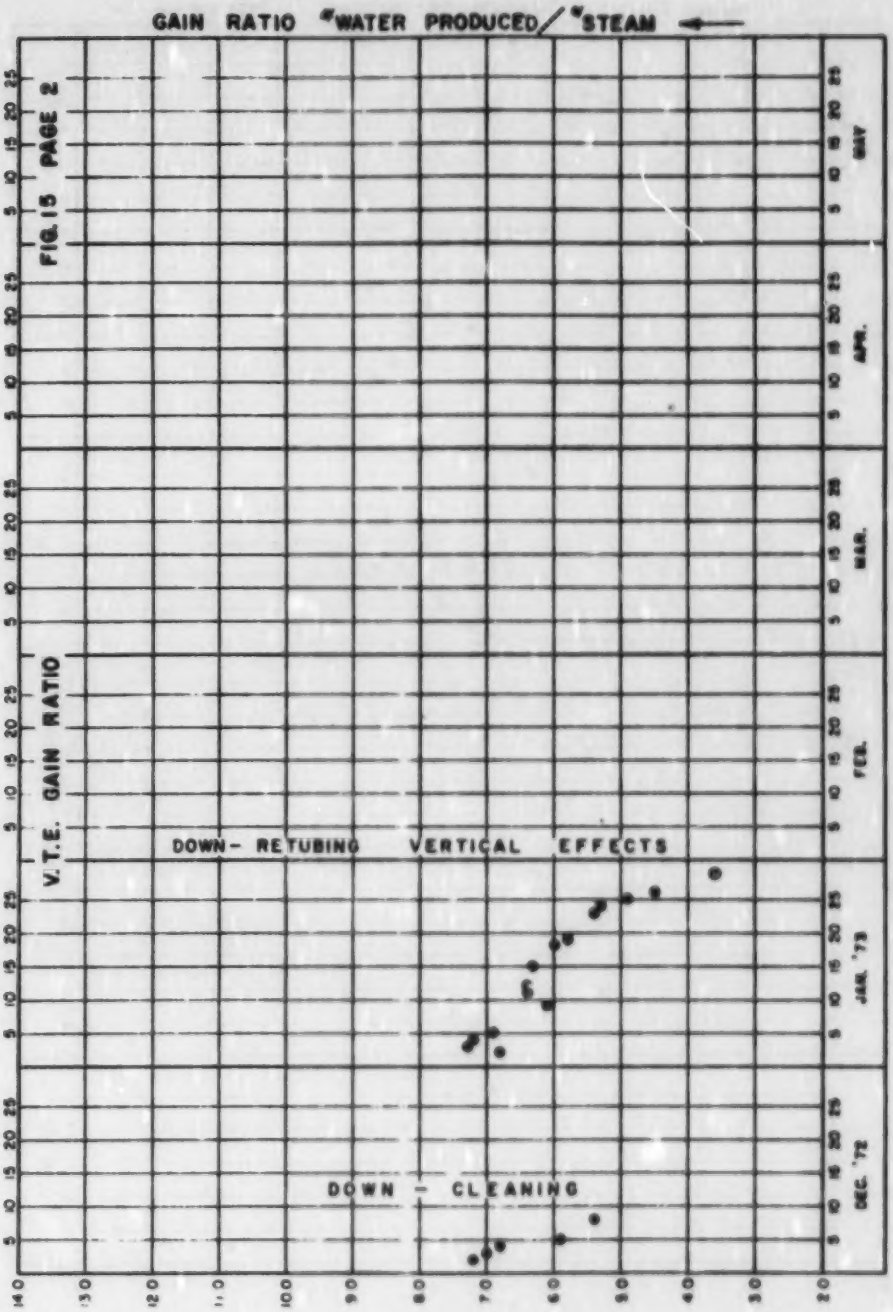


Figure 15 (continued). Gain Ratio vs. Time

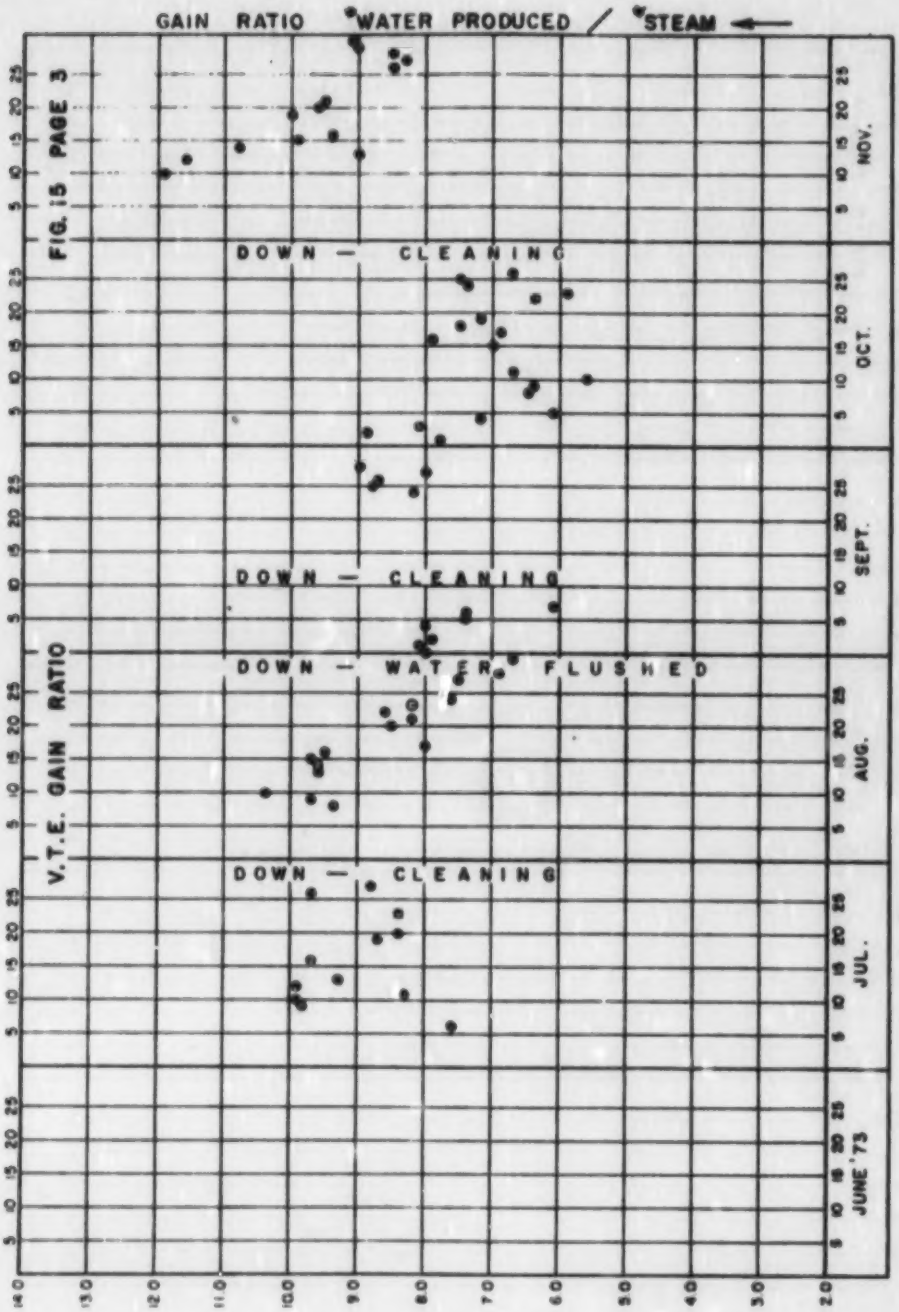


Figure 15 (continued). Gain Ratio vs. Time

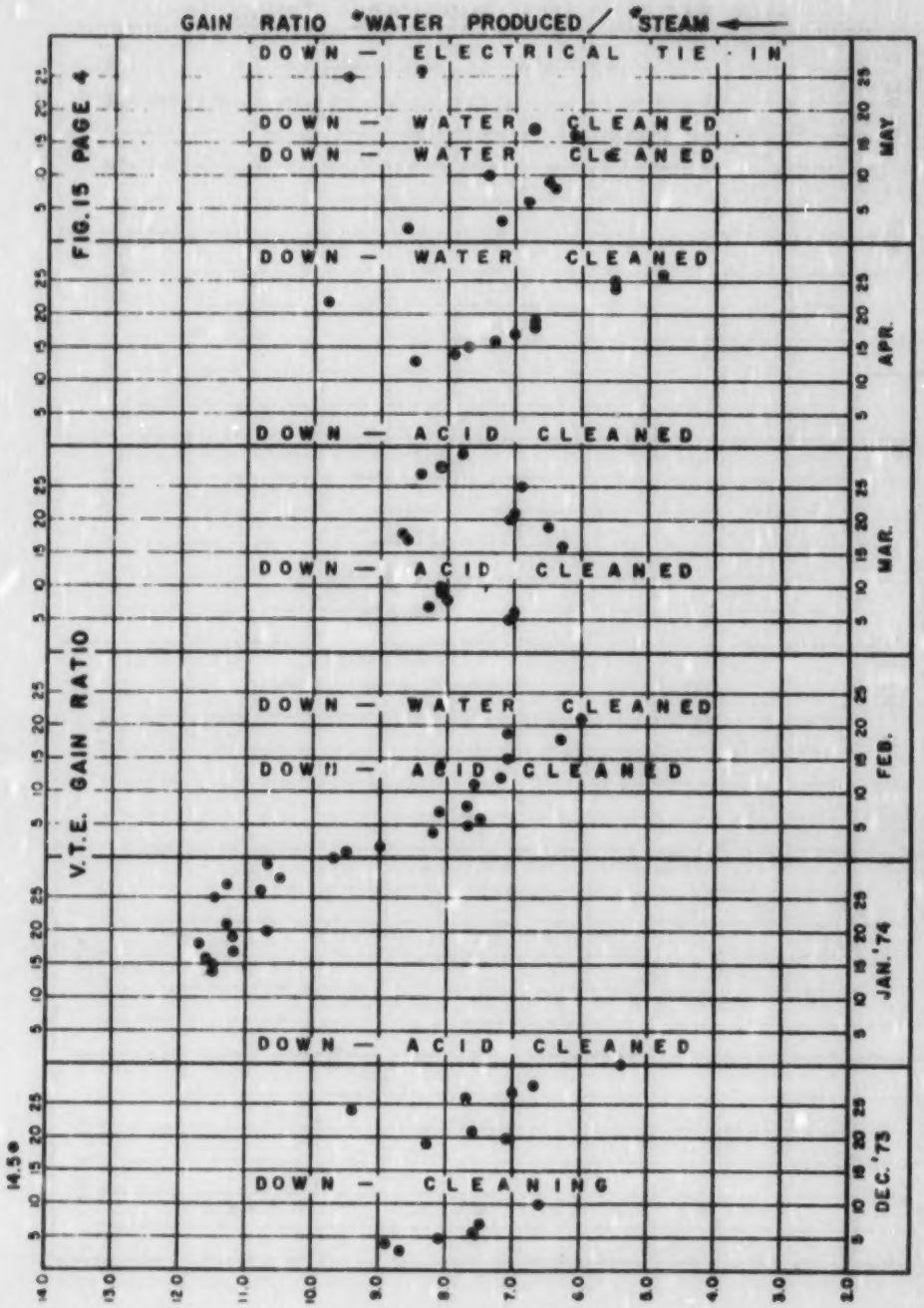
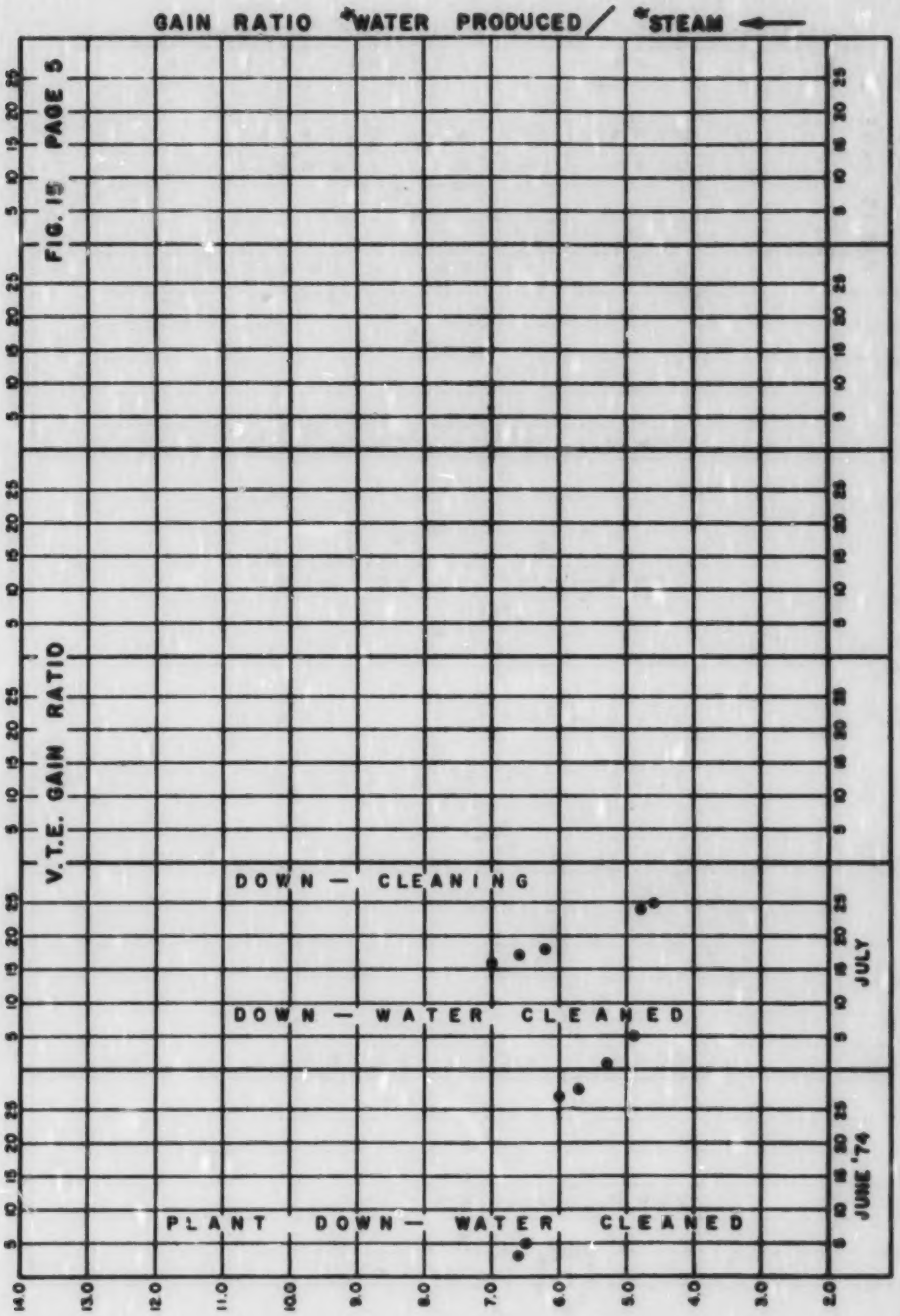


Figure 15 (continued). Gain Ratio vs. Time



ECONOMICS

CAPITAL COST (Refer to Table 12)

Purchase cost of the mechanical, battery-limits equipment for the VTE; including the evaporator, feedheaters, pumps, instruments, degassing and vacuum equipment, and supporting structural steel for these items; but excluding the control room, cooling tower, electrical gear, and installation; was approximately $550,000. If titanium tubing had been specified originally, the purchase cost of these items would have been $590,000.

Total direct installed cost of the VTE facility was $1,184,784. This figure is for all components directly associated with the VTE and includes no investment in supporting facilities, such as utilities or previously existing wastewater treatment equipment.

Any process to concentrate or recover water from an effluent containing water soluble salts such as is the case with synthetic rubber manufacture and many other chemical operations must make provision for an ultimate sink to receive these water soluble materials to prevent them from re-entering the environment. For this reason, we have included in Table 12 the cost of constructing the 89-acre lined solar evaporation ponds which were built in 1968. Including the cost of these ponds and the cost to retube the unit with titanium, the total capital investment was $2,029,791. In addition, another $83,190 was spent for pilot plant facilities.

Table 12. CAPITAL COSTS AND PILOT PLANT COSTS

I.	Capital Cost Items	
	A. VTE unit installed (1971)[a]	$1,184,783.90
	B. Solar ponds (1968)	749,051.13
	C. Retubing VTE with titanium tubes (1973)	95,956.15
	Total capital cost	$2,029,791.18
II.	Expense Items (Pilot Plant)[b]	
	A. Chlorinator	$ 2,170.10
	B. Trickling filter	18,367.59
	C. Activated carbon filter	7,733.96
	D. Simulator	17,320.63
	E. Ozonator	37,598.45
	Pilot plant expenditures for equipment and materials	$ 83,190.73

[a](1971) Indicates year in which unit was constructed.

[b]Numbers for pilot plant do not contain any operating costs.

OPERATING COST

Table 13 summarizes the direct operating costs; overhead, depreciation, taxes, insurance, etc., are not included. The figures shown are the totals for the calendar months indicated. Unit costs in dollars per thousand gallons of product water ranged from a low of $0.80 to a high of $8.44 with an average of $3.848 per thousand for the period of July 1973 to and including July 1974 or, if the last two months which were very high are excluded, an average cost of $3.154 is obtained for the 11-month period July 1, 1973, to May 30, 1974.

These costs include the cost of operating the pilot operations, cleaning, and maintenance but do not include the cost of pilot plant equipment or materials to construct.

Table 14 shows the operating hours, downtime hours and the percent operating time for various periods of attempted continuous runs for the period of September 22, 1973, to July 29, 1974. This data is not available for earlier runs because at that time the on-stream times were considered too short to be of any value.

If the abnormal rate of fouling could be eliminated, the indications are that the unit would operate unattended and would require a minimum of maintenance. The net result should be an operating cost within the reach of most plants requiring the recovery of water from effluents containing water soluble salts and/or the concentration of brines for ultimate disposal. As stated on Page 51, the fuel cost per 1,000 gallons of water recovered would be 83% of the cost of 1,000 pounds of steam if the gain ratio could be maintained at 10.0.

Table 15 shows the consumption of steam and electricity per 1,000 gallons of water recovered and the cost per month for the period of June 1972 to July 1974.

Table 13. OPERATING COST DATA[a]

Month	Operating cost	Materials	Labor	1000 gals. product	Cost/1000 gals. prod.
June 1972	$ 2,747	$ 1	$ 1,802	3,526.66	$1.29
July	3,086	11	787	4,864.87	.80
Aug.	2,834	176	3,422	6,700.14	.96
Sept.	3,301	4,270	10,038	4,380.43	4.02
Oct.	5,018	576	2,496	3,019.74	2.68
Nov.	5,994	957	21,702	3,968.84	7.22
Dec.	3,633	802	5,343	1,176.99	8.31
Jan. 1973	9,653	202	3,466	5,671.37	2.35
Feb.	0	138	4,792	0	-
Mar.	0	87	7,026	0	-
Apr.	0	83	8,444	0	-
May	0	510	5,494	0	-
June	381	608	5,397	0	-
July	10,618	159	4,115	4,346.40	3.42
Aug.	6,219	140	5,107	6,398.13	1.79
Sept.	6,653	868	5,304	4,485.50	2.86
Oct.	8,184	864	3,415	4,119.00	3.03
Nov.	7,245	2,438	3,860	5,322.10	2.54
Dec.	6,848	968	4,088	4,914.43	2.42
Jan. 1974	5,720	1,356	3,097	4,481.83	2.27
Feb.	6,990	1,538	3,204	3,639.06	3.22
Mar.	6,231	399	1,025	1,141.00	6.71
Apr.	4,608	1,121	2,708	2,657.68	3.17
May	7,485	1,786	2,730	3,684.27	3.26
June	9,203	607	2,131	1,732.61	6.89
July	14,748	968	2,578	2,167.22	8.44

[a] - See following page

Table 13 (continued). OPERATING COST DATA[a]

[a] - Definitions:

Operating costs - Includes only treatment chemicals, steam, electricity and cleaning chemicals used.

Materials - Maintenance supplies, seals, pumps, pipe, fittings, valves, etc., excluding equipment and materials used in the construction of the pilot plants.

Labor - All direct labor charged to the unit for operation and pilot plant work. Engineers, maintenance, operation including labor to construct and alter pilot plants.

Table 14. PERCENT VTE OPERATING TIME

Startup	Shutdown	Operating hrs.	Downtime hrs.	% Operating time
9/22/73(00:30)	10/27/73(04:00)	839	328	72%
11/9/73 (20:00)	12/10/73(14:00)	710	171	81
12/17/73(17:00)	1/2/74(04:00)	368	276	57
1/13/74(16:00)	2/12/74(04:00)	636	41	94
2/13/74(21:00)	2/22/74(21:30)	221	236	48
3/4/74 (01:00)	3/11/74(04:00)	173	110	61
3/15/74(18:00)	3/30/74(13:00)	303	296	51
4/11/74(21:00)	4/29/74(01:30)	404	72	85
5/2/74 (02:00)	5/15/74(04:00)	310	18	95
5/15/74(22:00)	5/20/74(06:00)	104	110	49
5/24/74(20:00)	6/9/74 (16:00)	304	384	79
6/25/74(16:00)	7/8/74 (02:00)	258	118	69
7/13/74(00:00)	7/29/74(03:00)	387		

Table 15. STEAM AND ELECTRICITY USAGE

Month	Lbs. Steam/ 1,000 Gals. Prod.	Steam Cost $/Month	Electricity KWH/1,000 Gals. Prod.	Elec. Cost $/Month	Total $/1,000 Gals.
June 1972	1,047.3	$2,597	44.8	$1,112	$1.05
July	653.0	2,240	20.5	700	.60
Aug.	397.6	1,878	15.0	725	.39
Sept.	664.3	2,052	29.4	926	.68
Oct.	1,837.0	3,911	38.0	827	1.57
Nov.	1,741.0	4,871	25.1	717	1.41
Dec.	3,516.6	2,918	66.8	566	2.95
Jan. 1973	2,080.8	8,320	23.0	927	1.63
Feb.	0	0	0	0	-
Mar.	0	0	0	0	-
Apr.	0	0	0	0	-
May	0	0	0	0	-
June	0	0	0	0	-
July	2,850.7	8,735	52.5	1,642	2.39
Aug.	1,128.4	5,090	15.7	754	.91
Sept.	1,646.1	5,206	23.0	754	1.33
Oct.	2,046.6	5,943	28.6	920	1.67
Nov.	2,007.0	5,648	23.5	779	1.21
Dec.	1,180.4	5,972	10.6	657	1.35
Jan. 1974	1,470.8	4,746	19.8	755	1.23
Feb.	2,314.1	6,063	24.7	763	1.88
Mar.	6,222.8	5,112	94.7	919	5.29
Apr.	1,901.7	3,639	31.6	713	1.64
May	2,382.0	6,319	29.4	922	1.97
June	6,637.0	8,280	43.6	642	5.15
July	2,168.7	3,450	48.3	869	1.99

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4. Bearden, D. F., "Activated Carbon Pilot Plant," Unpublished Interoffice Memo, The General Tire & Rubber Company, Odessa, Texas, June 12, 1973.

5. Koraido, D. L., "Activated Carbon Adsorption Pilot Study," Unpublished Letter - Calgon Corp. to The General Tire & Rubber Company, Odessa, Texas, May 14, 1973.

APPENDIX

Table 16. VTE OPERATING DATA NOMENCLATURE

Waste feed:	Clarified wastewater feed to the VTE, gallons per minute.
Product:	Pure product water produced by the VTE, including condensed steam fed to the first effect but not evaporated water sent to the cooling tower, gallons per minute.
Main steam:	Steam fed to first effect plus steam consumed by air ejectors, pounds per minute.
Eff. 1 stm.:	Steam fed to first effect, pounds per minute.
Prod. cond.:	Product conductivity, micromhos. Under "FLOWS," conductivity is the instantaneous reading taken from control panel indicator when daily data is logged. The product conductivity under "PRODUCT" is the value from laboratory analysis of the grab sample.
Gain ratio:	Plant efficiency measurement, pounds of product per pound of steam to first effect.
Brine ratio:	Brine ratio, quantity of brine divided by quantity of waste feed.
Sul.:	Sulfide, parts per million.
HX301:	Sample taken from preheated feed stream.
Deaerator:	Sample taken from Deaerator/Decarbonator vessel.

Table 16. VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972							1972
Flows	6/27	6/28	6/29	6/30	7/3	7/5	7/6	7/7
Waste feed gpm	323	320	323	321	315	267	270	265
Product gpm	221	220	207	191	206	195	178	65
Main stm. #/min.	178	177	-	175	185	167	153	155
Eff. 1 stm. #/min.	171	172	169	168	176	159	143	141
Prod. cond.	8.5	18	1	5	14	32	18	50+
Gain ratio	10.8	10.7	10.2	9.5	9.8	10.2	10.4	-
Brine ratio	0.316	0.313	0.359	0.405	0.346	0.270	.341	-
Feed								
pH								
Cl^-, ppm								
Sul., deaer.								
pH, HX301								
Product								
pH								
Prod. cond.		22	38	16	10	55	26	
Cl^-, ppm		7	6	7	1.4	3	3	
NH_3, ppm		4	6	3	2	4	6	
Blowdown								
pH								
Cl^-, ppm								

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972						1972
Flows		7/24	7/26	7/27	7/28	7/29	8/5
Waste Feed gpm	Down - Due to Ammonia Spill	313	305	311	339	227	303
Product gpm		175	178	182	202	135	199
Main stm. #/min.		174	170	180	194	130	182
Eff. 1 stm. #/min.		170	169	172	187	124	171
Prod. cond.		18	11	11	7	6.5	22
Gain ratio		8.6	8.8	8.8	9.0	9.1	9.7
Brine ratio		0.441	0.416	0.415	0.404	0.405	0.343
Feed							
pH							
Cl^-, ppm							
Sul., deaer.						0.00	
pH, HX301						5.1	
Product							
pH						6.9	
Prod. cond.		42	31	23	17	16	
Cl^-, ppm		7	4.4	6.7	5	7	
NH_3, ppm		10.0	4.0	1.5	3	2	
Blowdown							
pH						8.0	
Cl^-, ppm						704	

Down - Ammonia Spill

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972							1972
Flows	8/6	8/7	8/8	8/9	8/10	8/11	8/12	8/13
Waste feed gpm	313	311	300	308	302	300	301	300
Product gpm	192	192	133	174	212	206	215	206
Main stm. #/min.	183	183	199	197	150	200	198	198
Eff. 1 stm. #/min.	174	175	189	186	188	188	188	183
Prod. cond.	7	12	13	12	14	15	15	14
Gain ratio	9.2	9.2	5.9	7.8	9.4	9.1	9.5	9.4
Brine ratio	0.386	0.383	0.557	0.435	0.298	0.313	0.286	0.313
Feed								
pH	7.6	7.5	7.4	7.4	7.4	7.5	7.1	
Cl^-, ppm	516	563	721	475	448	246	105	
Sul., deaer.	0.10	0.00	0.02	0.01	0.05	0.01	0.01	
pH, HX301	6.3	7.6	7.4	7.1	6.8	7.0	6.3	
Product								
pH	7.4	8.2	8.6	8.4	8.3	8.1	8.1	
Prod. cond.	17	32	41	25	26	26	25	
Cl^-, ppm	3	5	4	6	8	5	5	
NH_3, ppm	5	6	7	5	5	6	6	
Blowdown								
pH	6.0	6.7			6.6	6.7	6.4	
Cl^-, ppm	1069	1178			1037	1195	1225	

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972							1972
Flows	8/14	8/15	8/16	8/17	8/18		8/30	8/31
Waste feed gpm	265	295	295	295	295	Down - Cleaning	307	330
Product gpm	203	195	199	195	192		172	175
Main stm. #/min.	198	192	200	198	197		172	177
Eff. 1 stm. #/min.	183	178	184	183	182		156	162
Prod. cond.	14	11	12	11	14		25	17
Gain ratio	9.3	9.1	9.0	8.9	8.8		9.2	9.0
Brine ratio	0.233	0.339	0.325	0.339	0.349		0.440	0.470
Feed								
pH		6.1	7.2	7.4	6.5		7.1	7.4
Cl^-, ppm		378	317	115	300		369	288
Sul., deaer.		0.03	0.09	0.02	0.05		0.09	0.12
pH, HX301		6.8	6.7	7.1	7.1		7.6	7.9
Product								
pH		8.9	9.0	9.2	9.0		8.2	8.2
Prod. cond.		16	-	-	23		28	19
Cl^-, ppm		4	3	2	4		3	3
NH_3, ppm		5	5	5	6		6	6
Blowdown								
pH		6.2	6.3	6.2	6.6		7.4	7.7
Cl^-, ppm		1055	967	987	703		879	615

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972							1972
Flows	9/1	9/5	9/6	9/7	9/8	9/11	9/12	
Waste feed gpm	321	332	323	323	286	240	256	Down - Cleaning
Product gpm	167	151	147	137	122	96	97	
Main stm. #/min.	172	173	171	168	163	146	147	
Eff. 1 stm. #/min.	142	159	159	153	147	131	131	
Prod. cond.	18	6	5	5	12	7	10	
Gain ratio	9.8	7.9	7.7	7.5	6.9	6.1	6.2	
Brine ratio	0.480	0.545	0.545	0.576	0.573	0.600	0.621	
Feed								
pH	7.2	6.3	6.2	8.4	8.0			
Cl^-, ppm	390	422	352	317	223			
Sul., deaer.	0.17	0.07	0.06	0.06	0.09			
pH, HX301	7.4	6.1	6.3	6.9	7.0			
Product								
pH	7.7	7.1	7.1	7.1	7.5			
Prod. cond.	19	10	9	11	15			
Cl^-, ppm	3	5	5	5	6			
NH_3, ppm	5	3	4	5	4			
Blowdown								
pH	7.1	6.2	5.9	5.8	6.7			
Cl^-, ppm	879	967	967	704	523			

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972 9/17	9/18	9/19	9/20	9/22	9/25		1972 9/28
Flows								
Waste feed gpm	332	414	384	295	295	203	Down - Duo-Vac Filter Repairs	301
Product gpm	173	143	163	168	132	116		159
Main stm. #/min.	164	149	177	171	153	147		211
Eff. 1 stm. #/min.	153	143	164	156	152	134		195
Prod. cond.	15	15	9	8	15	12		6
Gain ratio	9.4	8.3	8.3	9.0	7.2	7.2		6.8
Brine ratio	0.479	0.655	0.576	0.431	0.553	0.429		0.472
Feed								
pH		6.7	6.2	6.4	6.3			
Cl^-, ppm		528	564	440	492			
Sul., deaer.		0.07	0.05	0.03	0.40			
pH, HX301		6.0	6.6	6.9	7.0			
Product								
pH		7.5	7.2	7.2	7.5			
Prod. cond.		14	12	11	15			
Cl^-, ppm		3	4	4	6			
NH_3, ppm		3	2	3	5			
Blowdown								
pH		6.8	6.3	6.7	6.5			
Cl^-, ppm		967	940	985	1126			

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972						1972
Flows		10/2	10/3	10/4	10/5	10/6	10/9	10/10
Waste feed gpm	Down - Alum Leak on Coag	347	352	250	257	255	355	354
Product gpm		167	178	136	151	151	158	152
Main stm. #/min.		225	220	163	180	182	212	206
Eff. 1 stm. #/min.		211	200	141	160	163	177	172
Prod. cond.		8	2.5	7	8	8	10	14
Gain ratio		6.6	7.4	8.0	7.9	7.7	7.4	7.4
Brine ratio		0.519	0.494	0.456	0.412	0.408	0.555	0.571
Feed								
pH			5.2	6.1	6.2	5.2		6.6
Cl^-, ppm			466	383	357	308		440
Sul., deaer.			0.38	0.14	0.12	0.08		0.14
pH, HX301			6.0	6.8	7.2	7.0		6.9
Product								
pH			7.5	8.2	8.4	7.7		7.4
Prod. cond.			6	10	10	10		17
Cl^-, ppm			6	8	6	8		6
NH_3, ppm			3	3	3	4		5
Blowdown								
pH			6.1	6.8	6.9	6.6		6.8
Cl^-, ppm			967	957	1020	915		932

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972						1972	
Flows	10/11	10/12	10/13	10/16	10/17		11/3	11/6
Waste feed gpm	355	362	319	251	258	Down - Cleaning, Replace Leaking Feedheater Tubes	298	303
Product gpm	154	149	138	117	132		131	116
Main stm. #/min.	210	209	201	173	183		188	187
Eff. 1 stm. #/min.	176	173	167	141	149		165	167
Prod. cond.	12	13	16	18	20		11	7.5
Gain ratio	7.3	7.2	6.9	6.9	7.4		6.6	5.8
Brine ratio	0.566	0.588	0.567	0.534	0.488		0.560	0.617
Feed								
pH	6.1	5.6	5.3		6.6			
Cl^-, ppm	458	598	572		633			
Sul., deaer.	0.15	0.22	0.23		0.42			
pH, HX301	6.7	6.5	6.8		7.1			
Product								
pH	7.2	6.8	7.6		8.1			
Prod. cond.	17	18	21		26			
Cl^-, ppm	7	9	7		8			
NH_3, ppm	7	5	6		6			
Blowdown								
pH	6.5	6.3	6.5		6.6			
Cl^-, ppm	793	1196	1073		1337			

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972							1972
Flows	11/7	11/8	11/9	11/11	11/12	11/13	11/14	11/15
Waste feed gpm	278	285	273	265	265	261	254	232
Product gpm	143	142	146	146	142	144	139	120
Main stm. #/min.	190	201	203	208	203	206	203	188
Eff. 1 stm. #/min.	173	177	183	185	183	185	183	164
Prod. cond.	5	3.5	6	13	15	12	13	14
Gain ratio	6.9	6.7	6.7	6.6	6.5	6.5	6.3	6.1
Brine ratio	0.486	0.502	0.465	0.449	0.464	0.448	0.453	0.483
Feed								
pH	8.4	8.8	8.8			8.7	8.7	
Cl^-, ppm	434	438	398			438	390	
Sul., deaer.	0.10	0.07	0.05			0.12	0.15	
pH, HX301	7.4	6.3	6.8			6.7	7.4	
Product								
pH	6.3	6.3	6.8			7.9	8.4	
Prod. cond.	8	8	10			16	17	
Cl^-, ppm	7	8	7			11	11	
NH_3, ppm	3	3	3			5	5	
Blowdown								
pH	6.0	6.1	6.5			6.7	7.2	
Cl^-, ppm	869	883	801			1175	727	

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972 11/18	11/20	11/21	11/22	11/25	11/26	1972 11/27
Flows							
Waste feed gpm	267	263	264	262	263	269	263
Product gpm	146	133	138	140	119	124	122
Main stm. #/min.	214	214	219	227	238	236	234
Eff. 1 stm. #/min.	191	189	196	222	197	199	195
Prod. cond.	12	4	4.5	4.5	4.5	4	4
Gain ratio	6.4	5.9	5.9	5.3	5.0	5.2	5.2
Brine ratio	0.453	0.494	0.477	0.466	0.548	0.539	0.536
Feed							
pH			8.3	8.6	8.5	8.8	8.5
Cl^-, ppm			372	351	160	369	365
Sul., deaer.			0.02	0.02	0.14	0.05	0.09
pH, HX301			6.4	5.5	7.6	7.8	6.7
Product							
pH			6.9	5.8	7.0	7.1	6.2
Prod. cond.			8	8	12	6	7
Cl^-, ppm			8	7	5	4	5
NH_3, ppm			4	4	4	4	4
Blowdown							
pH			6.2	5.4	7.3	7.5	7.3
Cl^-, ppm			766	715	642	651	691

Down - Low Effluent Pond Level

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1972							1972
Flows	11/28		12/2	12/3	12/4	12/5	12/8	
Waste feed gpm	258	Down - Cleaning	226	228	227	213	192	Down - Cleaning
Product gpm	108		151	147	146	119	117	
Main stm. #/min.	230		211	213	224	211	224	
Eff. 1 stm. #/min.	182		175	175	180	167	182	
Prod. cond.	4		10	7	6	4	20	
Gain ratio	4.9		7.2	7.0	6.8	5.9	5.4	
Brine ratio	0.581		0.332	0.355	0.357	0.441	0.391	
Feed								
pH	8.4			8.5	7.7			
Cl^-, ppm	383			539	504			
Sul., deaer.	0.11			0.04	0.03			
pH, HX301	6.8			6.8	5.7			
Product								
pH	6.1			6.2	6.9			
Prod. cond.	6			10	155			
Cl^-, ppm	4			5	10			
NH_3, ppm	3			7	8			
Blowdown								
pH	6.5			6.1	5.6			
Cl^-, ppm	700			1649	1596			

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973							1973
	1/2	1/3	1/4	1/5	1/9	1/11	1/12	1/15
Flows								
Waste feed gpm	301	327	326	343	343	230	292	271
Product gpm	146	161	142	187	157	133	161	149
Main stm. #/min.	211	236	213	268	264	227	265	246
Eff. 1 stm. #/min.	161	184	165	227	216	173	211	196
Prod. cond.	13	15	11	12	20	23	27	29
Gain ratio	6.8	7.3	7.2	6.9	6.1	6.4	6.4	6.3
Brine ratio	0.515	0.508	0.564	0.455	0.542	0.422	0.449	0.450
Feed								
pH		8.1	7.8	7.5			7.0	
Cl^-, ppm		399	381	301			393	
Sul., deaer.		0.02	0.07	0.08			0.07	
pH, HX301		7.4	6.4	6.9			6.9	
Product								
pH		7.6	6.8	6.8			8.4	
Prod. cond.		11	7	12			13	
Cl^-, ppm		3	3	3			2	
NH_3, ppm		4	3	4			3	
Blowdown								
pH		7.2	6.5	6.7			7.0	
Cl^-, ppm		780	709	691			1000	

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973 1/18	1/19	1/23	1/24	1/25	1/26	1/29	1973
Flows								
Waste feed gpm	292	293	289	309	291	310	255	Down - Retubing Vertical Effects 1-8
Product gpm	148	143	132	136	115	116	78	
Main stm. #/min.	250	245	265	275	256	266	286	
Eff. 1 stm. #/min.	206	204	202	213	197	215	179	
Prod. cond.	25	23	13	11	11	11	21	
Gain ratio	6.0	5.8	5.4	5.3	4.9	4.5	3.6	
Brine ratio	0.493	0.512	0.543	0.560	0.605	0.626	0.694	
Feed								
pH					6.7	6.7		
Cl^-, ppm					426	400		
Sul., deaer.					0.08	0.12		
pH, HX301					7.4	7.1		
Product								
pH					8.3	8.7		
Prod. cond.					14	13		
Cl^-, ppm					4	4		
NH_3, ppm					5	5		
Blowdown								
pH					6.9	6.6		
Cl^-, ppm					775	746		

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973							1973
Flows	7/6	7/9	7/10	7/11	7/12	7/13	7/16	7/19
Waste feed gpm	265	316	314	315	319	314	264	299
Product gpm	209	230	215	223	219	225	181	198
Main stm. #/min.	203	215	204	222	213	219	189	213
Eff. 1 stm. #/min.	229	195	182	224	184	202	156	190
Prod. cond.	-	-	17	20	16	16	9	11
Gain ratio	7.6	9.8	9.9	8.3	9.9	9.3	9.7	8.7
Brine ratio	0.211	0.272	0.315	0.292	0.313	0.283	0.314	0.338

Parameter	1973							1973
Flows	7/20	7/23	7/26	7/27		8/8	8/9	8/10
Waste feed gpm	326	307	273	243	Down - Cleaning	304	304	301
Product gpm	195	174	130	113		215	206	215
Main stm. #/min.	215	204	155	146		207	200	197
Eff. 1 stm. #/min.	193	173	112	107		191	177	172
Prod. cond.	11	16	14	12		17	16	17
Gain ratio	8.4	8.4	9.7	8.8		9.4	9.7	10.4
Brine ratio	0.402	0.433	0.524	0.535		0.293	0.322	0.286

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973							1973
Flows	8/13	8/14	8/15	8/16	8/17	8/20	8/21	8/22
Waste feed gpm	301	247	248	249	333	273	271	246
Product gpm	203	185	188	186	201	186	183	178
Main stm. #/min.	201	188	192	192	213	203	202	199
Eff. 1 stm. #/min.	176	160	162	164	209	183	185	173
Prod. cond.	15	13	14	15	16	17	13	17
Gain ratio	9.6	9.6	9.7	9.5	8.0	8.5	8.2	8.6
Brine ratio	0.326	0.251	0.242	0.253	0.396	0.319	0.325	0.276

Parameter	1973							1973
Flows	8/23	8/24	8/27	8/28		8/30	8/31	9/1
Waste feed gpm	246	274	227	227	Down - Flushed with City Water	227	225	226
Product gpm	175	182	152	145		148	159	165
Main stm. #/min.	203	211	197	202		209	205	209
Eff. 1 stm. #/min.	177	200	170	176		184	166	170
Prod. cond.	17	19	18	19		19	5	16
Gain ratio	8.2	7.6	7.5	6.9		6.7	8.0	8.1
Brine ratio	0.289	0.336	0.330	0.361		0.348	0.293	0.270

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973 9/2	9/4	9/5	9/6	9/7		1973 9/24	9/25
Flows								
Waste feed gpm	277	225	223	212	293	Down - Cleaning	354	302
Product gpm	162	159	156	149	153		217	198
Main stm. #/min.	207	204	209	207	226		220	200
Eff. 1 stm. #/min.	170	166	176	169	209		220	187
Prod. cond.	14	14	14	17	17		14	11
Gain ratio	7.9	8.0	7.4	7.4	6.1		8.2	8.8
Brine ratio	0.415	0.293	0.300	0.297	0.478		0.387	0.344

Parameter	1973 9/26	9/27	9/28	10/1	10/2	10/3	1973 10/4	10/5
Flows								
Waste feed gpm	301	184	283	286	208	205	203	258
Product gpm	203	111	188	185	157	163	159	162
Main stm. #/min.	201	123	193	203	181	196	139	225
Eff. 1 stm. #/min.	194	115	174	197	147	168	185	223
Prod. cond.	10	22	20	11	11	17	11	13
Gain ratio	8.7	8.0	9.0	7.8	8.9	8.1	7.2	6.1
Brine ratio	0.326	0.397	0.336	0.353	0.245	0.205	0.217	0.372

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973							1973
Flows	10/8	10/9	10/10	10/11	10/15	10/16	10/17	10/18
Waste feed gpm	299	326	325	313	301	295	264	249
Product gpm	161	171	172	170	170	161	159	168
Main stm. #/min.	239	252	258	250	245	234	227	223
Eff. 1 stm. #/min.	206	224	255	213	202	170	191	187
Prod. cond.	12	12	12	13	-	11	-	12
Gain ratio	6.5	6.4	5.6	6.7	7.0	7.9	6.9	7.5
Brine ratio	0.462	0.475	0.471	0.457	0.435	0.454	0.398	0.325

Parameter	1973							1973
Flows	10/19	10/22	10/23	10/24	10/25	10/26		11/10
Waste feed gpm	257	281	299	273	255	295	Down - Cleaning	278
Product gpm	157	144	142	150	169	162		213
Main stm. #/min.	223	228	241	222	233	243		-
Eff. 1 stm. #/min.	182	189	201	168	189	203		149
Prod. cond.	16	6	7	14	-	11		-
Gain ratio	7.2	6.4	5.9	7.4	7.5	6.7		11.9
Brine ratio	0.389	0.488	0.525	0.451	0.337	0.451		0.234

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973							1973
Flows	11/12	11/13	11/14	11/15	11/16	11/19	11/20	11/21
Waste feed gpm	307	263	303	307	303	316	327	286
Product gpm	226	160	220	204	194	194	195	166
Main stm. #/min.	219	192	223	227	232	238	250	219
Eff. 1 stm. #/min.	162	148	170	172	172	162	169	145
Prod. cond.	15	24	-	18	18	13	13.5	17
Gain ratio	11.6	9.0	10.8	9.9	9.4	10.0	9.6	9.5
Brine ratio	0.264	0.392	0.274	0.336	0.360	0.386	0.404	0.421

Parameter	1973							1973
Flows	11/26	11/27	11/28	11/29	11/30	12/3	12/4	12/5
Waste feed gpm	298	287	282	285	302	302	243	253
Product gpm	174	162	162	166	170	164	152	141
Main stm. #/min.	247	239	235	229	229	229	210	211
Eff. 1 stm. #/min.	170	163	159	154	155	157	143	145
Prod. cond.	14	17	15	16	11	18	11	6
Gain ratio	8.5	8.3	8.5	9.0	9.1	8.7	8.9	8.1
Brine ratio	0.416	0.436	0.426	0.418	0.437	0.457	0.374	0.443

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1973						1973	
Flows	12/6	12/7	12/10		12/18	12/19	12/20	12/21
Waste feed gpm	243	249	371	Down - Cleaning	332	272	279	298
Product gpm	132	128	123		228	181	143	183
Main stm. #/min.	214	215	233		245	234	218	254
Eff. 1 stm. #/min.	144	143	155		131	182	167	202
Prod. cond.	8.5	11	31		11	9	9	-
Gain ratio	7.6	7.5	6.6		14.5	8.3	7.1	7.6
Brine ratio	0.457	0.486	0.668		0.313	0.335	0.487	0.386

Parameter	1973						1974	
Flows	12/24	12/26	12/27	12/28	12/31		1/14	1/15
Waste feed gpm	203	333	291	296	336	Down - Acid Cleaned	345	323
Product gpm	138	170	152	142	136		229	226
Main stm. #/min.	184	256	232	226	271		237	238
Eff. 1 stm. #/min.	123	185	182	178	209		166	164
Prod. cond.	15	19	17	18	16		7	6
Gain ratio	9.4	7.7	7.0	6.7	5.4		11.5	11.5
Brine ratio	0.320	0.489	0.478	0.520	0.595		0.336	0.300

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1974							1974
Flows	1/16	1/17	1/18	1/19	1/20	1/21	1/25	1/26
Waste feed gpm	293	287	286	310	283	256	238	200
Product gpm	213	193	192	190	183	173	158	150
Main stm. #/min.	229	215	204	210	211	197	181	170
Eff. 1 stm. #/min.	154	144	137	142	143	128	115	116
Prod cond.	5	3	4	-	-	3	21	-
Gain ratio	11.6	11.2	11.7	11.2	10.7	11.3	11.5	10.8
Brine ratio	0.273	0.328	0.329	0.387	0.353	0.324	0.336	0.250

Parameter	1974							1974
Flows	1/27	1/28	1/30	1/31	2/1	2/2	2/4	2/5
Waste feed gpm	244	305	308	345	327	250	282	290
Product gpm	185	203	202	200	170	150	168	171
Main stm. #/min.	202	237	232	238	214	195	248	255
Eff. 1 stm. #/min.	137	161	158	172	149	139	170	185
Prod. cond.	-	7	2	3	-	-	6	5
Gain ratio	11.3	10.5	10.7	9.7	9.5	9.0	8.2	7.7
Brine ratio	0.242	0.334	0.344	0.420	0.480	0.400	0.404	0.410

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1974						1974	
Flows	2/6	2/7	2/8	2/11	2/12		2/14	2/15
Waste feed gpm	321	309	358	309	383	Down - Acid Cleaned	275	345
Product gpm	182	172	183	157	188		153	172
Main stm. #/min.	285	261	286	247	304		219	269
Eff. 1 stm. #/min.	203	177	198	172	219		158	201
Prod. cond.	3	2	1	-	-		-	14
Gain ratio	7.5	8.1	7.7	7.6	7.2		8.1	7.1
Brine ratio	0.433	0.443	0.489	0.492	0.509		0.444	0.501

Parameter	1974							1974
Flows	2/18	2/19	2/21		3/5	3/6	3/7	3/8
Waste feed gpm	313	289	313	Down - Water Cleaned	261	306	303	331
Product gpm	147	157	150		147	171	192	190
Main stm. #/min.	274	264	293		213	251	236	242
Eff. 1 stm. #/min.	196	185	209		173	203	193	198
Prod. cond.	15	23	7		9	10	-	9
Gain ratio	6.3	7.1	6.0		7.1	7.0	8.3	8.0
Brine ratio	0.530	0.457	0.521		0.437	0.441	0.366	0.426

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1974							1974
Flows	3/9	3/10		3/16	3/17	3/18	3/19	3/20
Waste feed gpm	362	344	Down - Acid Cleaned	306	382	275	282	270
Product gpm	204	194		198	183	188	145	152
Main stm. #/min.	257	243		238	218	229	229	254
Eff. 1 stm. #/min.	211	200		264	178	180	185	179
Prod. cond.	-	-		-	-	16	23	17
Gain ratio	8.1	8.1		6.3	8.6	8.7	6.5	7.1
Brine ratio	0.436	0.436		0.353	0.521	0.316	0.486	0.437

Parameter	1974							1974
Flows	3/21	3/25	3/27	3/28	3/30		4/13	4/14
Waste feed gpm	278	289	290	302	291	Down - Acid Cleaned	249	248
Product gpm	160	153	195	195	172		178	164
Main stm. #/min.	235	227	237	244	226		200	200
Eff. 1 stm. #/min.	191	185	193	200	185		175	174
Prod. cond.	20	24	10	11	7		14	9
Gain ratio	7.0	6.9	8.4	8.1	7.8		8.5	7.9
Brine ratio	0.424	0.471	0.328	0.354	0.409		0.285	0.339

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1974							1974
Flows	4/15	4/16	4/17	4/18	4/19	4/22	4/24	4/25
Waste feed gpm	247	248	251	301	299	365	292	334
Product gpm	160	152	147	160	160	180	117	142
Main stm. #/min.	197	197	201	215	213	250	188	241
Eff. 1 stm. #/min.	173	174	174	200	198	153	179	215
Prod. cond.	18	17	15	12	14	17	14	14
Gain ratio	7.7	7.3	7.0	6.7	6.7	9.8	5.5	5.5
Brine ratio	0.352	0.387	0.414	0.468	0.465	0.509	0.599	0.575

Parameter	1974							1974
Flows	4/26		5/2	5/3	5/6	5/8	5/9	5/10
Waste feed gpm	312	Down - Water Cleaned	319	302	286	304	299	298
Product gpm	122		183	150	155	161	157	152
Main stm. #/min.	241		121	111	215	235	228	226
Eff. 1 stm. #/min.	212		177	173	191	211	203	171
Prod. cond.	19		24	12	13	13	13	13
Gain ratio	4.8		8.6	7.2	6.8	6.4	6.5	7.4
Brine ratio	0.609		0.426	0.503	0.458	0.470	0.475	0.490

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1974					1974		
Flows	5/13		5/16	5/17		5/25	5/26	
Waste feed gpm	300	Down - Water Cleaned	295	294	Down - Water Cleaned	306	300	Down - Electrical Tie-ins
Product gpm	152		165	155		187	168	
Main stm. #/min.	245		215	223		274	272	
Eff. 1 stm. #/min.	226		225	193		164	166	
Prod. cond.	13		19	16		-	-	
Gain ratio	5.6		6.1	6.7		9.5	8.4	
Brine ratio	0.493		0.441	0.473		0.389	0.440	

Parameter	1974							1974
Flows	6/3	6/5		6/27	6/28	7/1	7/5	
Waste feed gpm	310	299	Plant Down - Water Cleaned	308	269	269	271	Down - Water Cleaned
Product gpm	163	165		167	138	138	128	
Main stm. #/min.	298	313		352	504	337	591	
Eff. 1 stm. #/min.	205	211		234	201	219	219	
Prod. cond.	14	11		18	19	24	24	
Gain ratio	6.6	6.5		6.0	5.7	5.3	4.9	
Brine ratio	0.474	0.448		0.458	0.487	0.487	0.528	

Table 16 (continued). VTE OPERATING DATA, JUNE 27, 1972, TO JULY 25, 1974

Parameter	1974					1974
Flows	7/16	7/17	7/18	7/24	7/25	
Waste feed gpm	207	215	217	293	287	Down - Cleaning
Product gpm	132	135	133	120	113	
Main stm. #/min.	574	464	536	482	480	
Eff. 1 stm. #/min.	157	171	179	209	205	
Prod. cond.	18	17	20	8	8	
Gain ratio	7.0	6.6	6.2	4.8	4.6	
Brine ratio	0.362	0.372	0.387	0.590	0.606	